PB-212 960

EVALUATION OF THE FLUIDIZED BED COMBUSTION PROCESS. VOLUME II, TECHNICAL EVALUATION

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Westinghouse Research Laboratory

Prepared for:

Environmental Protection Agency

November 1971

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APTD-1166

PB 212 960

EVALUATION OF THE FLUIDIZED BED COMBUSTION PROCESS

Submitted to:

Office of Air Programs
Environmental Protection Agency

Contract No. CPA 70-9

By:

Westinghouse Research Laboratories
Pittsburgh, Pennsylvania

Reproduced by
NATIONAL TECHNICAL
INFORMATION SERVICE
U S Department of Commerce
Springfield VA 22151

Volume II TECHNICAL EVALUATION

61

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BIBLIOGRAPHIC DATA SHEET

1. Report No. APTD-1166

2.

3. Recipient's Accession No. PB 212 160

4. Title and Subtitle
Evaluation of the Fluidized Bed Combustion Process
Volume II Technical Evaluation

5. Report Date November 1971

6.

7. Author(s) D. H. Archer, D. L. Keairns, J. R. Hamm, et al

8. Performing Organization Rept. No.

9. Performing Organization Name and Address
Westinghouse Research Laboratories
Pittsburgh, Pennsylvania 15235

10. Project/Task/Work Unit No.

11. Contract/Grant No. CPA 70-9

12. Sponsoring Organization Name and Address
ENVIRONMENTAL PROTECTION AGENCY
Office of Air Programs
Research Triangle Park, North Carolina 27711

13. Type of Report & Period Covered

14.

15. Supplementary Notes

16. Abstracts This volume (Volume II) contains an evaluation of available data, an evaluation of fluidized bed combustion concepts for steam/power generation, a comparison of alternative concepts, conclusions and recommendations. Specifications and design concepts were established for industrial and utility fluidized bed boilers. Preliminary designs, performance projections, and cost estimates were prepared on the basis of the specifications and compared with conventional systems. The conventional power plant with combustion modification for NO_x control and wet scrubbers for SO_2 and particulate control is the most highly developed technology. Pressurized fluidized bed combustion technology is the most promising of its kind from the point of view of effective pollution abatement, efficient fuel utilization, and economical power generation. Atmospheric fluidized bed combustion may be better than conventional systems, but further development is required. The preliminary investigation coupled with previous experimental work of others points out the general feasibility of oil gasification/desulfurization as a retrofit SO_2 control system for utility boilers. Three advanced fluidized bed power generation systems and an industrial fluidized bed steam generation system were studied. Experimental development and systems evaluation were carried out for all three. The potential for pollution abatement, ability to use readily available fuels, promise for economical steam/power generation, present state of development, technical problem areas, advance design potential, and projected development time and cost were assessed for each system and compared with conventional fossil fuel-fired steam/power generating plants utilizing present pollution abatement technology.

17. Key Words and Document Analysis. 17a. Descriptors

Air pollution
Combustion
Boilers
Design criteria
Steam electric power generation
Expenses
Pressure
Feasibility
Fuel oils
Sulfur dioxide
Air pollution control equipment

17b. Identifiers/Open-Ended Terms

ia

17c. COSATI Field/Group 13B

18. Availability Statement Unlimited

19. Security Class (This Report) UNCLASSIFIED

20. Security Class (This Page) UNCLASSIFIED

21. No. of Pages 606 619

22. Price 9.00 0.95

PREFACE

The Office of Air Programs (OAP), of the United States Environmental Protection Agency, has organized and is sponsoring a fluidized bed fuel processing program. Its purpose is to develop and demonstrate new methods for utilizing fossil fuels -- particularly coal and oil -- in industrial boilers and in utility power plants. These methods should:

- Meet air pollution abatement goals for SO_2, NO_x, ash and smoke emissions
- Compete economically with alternative means for meeting these abatement goals.

Westinghouse has been assigned process evaluation tasks in the fluidized bed fuel processing program. These tasks have included:

- Conducting studies predicting fuel availability and usage costs; market surveys forecasting industrial boiler and utility power plant installations; and analyses projecting air pollution control improvement. The purpose of these investigations was to provide a basis for boiler design specifications -- capacities, steam conditions, pollutant emission restrictions, etc. -- and to determine the possible improvements in environmental and economic impact of various improved steam/power generation processes.
- Designing an industrial fluidized bed boiler and two utility boilers -- one carrying out combustion at one atm, the other at 10 atm pressure. All of the boilers use a limestone/dolomite sorbent in the fluidized bed for sulfur removal. Systems for regenerating and recycling the sorbent and for recovering the sulfur in useful form have also been

designed. All necessary auxiliaries for the industrial boiler and complete power plant equipment for the utility boilers have also been specified so that a complete system can be visualized. The purpose of these boiler system designs is to evaluate the performance and costs of industrial boilers and utility power plants employing fluidized bed combustion. Promising areas for future development have been identified. Westinghouse has been assisted in this work by Erie City Energy, Foster Wheeler, and United Engineers. The design and evaluation process has also pointed up technical areas where more information is needed and further development is required to meet the goal of a non-polluting, economical means of producing steam/power from fossil fuels.

- Conceptualizing a fluidized bed combustion boiler development plant. Preliminary designs have been produced for a development plant which will make possible an attack on the remaining technical problems in those areas where fluidized combustion has greatest potential for reducing air pollution and for economical generation of steam/power. Preliminary estimates of the cost for such a plant have also been made.

- Providing technical consultation and assistance on the OAP fluidized bed fuel processing program, including both combustion and gasification processes. Technical and economic comparisons have been carried out on various fluidized bed fuel processing systems and various conventional means of steam/power generation. An atmospheric pressure oil gasification/desulfurization process has been assessed in detail as an adjunct to a utility boiler.

The results of the various surveys, designs, evaluations, and comparisons are presented in this three-volume report. The report identifies fluidized bed fuel processing systems which should meet both market requirements and air pollution abatement requirements and which are likely to be cheaper than alternative, conventional systems. Finally, it recommends a program for commercializing promising processes.

This volume (Volume II) contains an evaluation of available data, an evaluation of fluidized bed combustion concepts for steam/power generation, a comparison of alternative concepts, conclusions and recommendations. Volume I is a summary of the complete report, and Volume III contains the detail surveys, design reports, and support studies.

ABSTRACT

The effectiveness and economics of fluidized bed combustion boilers in pollution abatement and steam/power generation have been evaluated. A 250,000 lb/hr coal-fired, factory-fabricated, industrial boiler has been designed along with all of its auxiliaries. Utility boilers have been designed for operation at atmospheric pressure in conventional steam power plants and for operation at 10 atm in a combined steam and gas turbine power plant. Overall capital and operating costs have been estimated for both 300 and 600 megawatt plants.

Of all the systems studied, the pressurized fluidized bed combustion boiler operating in a combined cycle power plant appears most effective in meeting projected emission standards and in reducing total SO_2, NO_x, and particulate emissions, and most economical in power generation. Such a plant has an estimated capital cost 20% less and energy cost 10% less than a conventional utility plant with a stack gas scrubber; this plant also has the greatest potential for increased generation efficiency and reduced costs.

A fluidized bed oil gasification-desulfurization system has also been designed and evaluated as an add-on unit for reducing SO_2 emissions from utility boilers burning high-sulfur oils. The estimated capital cost of such a unit is as much as 50% less than an add-on wet scrubbing system.

It is recommended that a developmental pressurized fluidized bed boiler unit of 10 to 30 megawatt capacity be designed and constructed to provide necessary technical information for a prototype power plant.

It is also recommended that the installation of a 50 to 200 megawatt demonstration fluidized bed oil gasifier-desulfurizer at a utility site be pursued.

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1. ASSESSMENT OF FLUIDIZED BED COMBUSTION

Specifications and design concepts were established for industrial and utility fluidized bed boilers. Preliminary designs, performance projections, and cost estimates were prepared on the basis of the specifications and compared with conventional systems.

The design specifications for industrial and utility fluidized bed combustion boilers are based on

- Preliminary surveys - market requirements, environmental control requirements
- Fluidized bed combustion experimental data and related technology experience.

PRELIMINARY SURVEYS

In order to carry out boiler designs and evaluations, it was necessary to gather market information on boiler applications and technical information on fluidized bed combustion. Market and other background data were collected for three general purposes:

- To establish functional specifications as a basis for boiler (or gasifier) designs. Boiler capacity, steam conditions, efficiency, operational features such as percent turn-down, and dynamic response are all established by customer requirements. In addition, a fuel is chosen on the basis of availability, cost, and convenience. Finally, government regulations establish maximum levels of pollutant (SO_2, NO_x, and particulates) emissions for boilers employing the selected fuel.
- To establish a basis of comparison. The performance and cost of the fluidized bed combustion boiler systems must be compared with alternative systems in order to evaluate the likelihood of their use.
- To establish the impact of a successful fluidized combustion boiler on air quality and on economical steam/power generation Fluidized bed combustion is a technology appropriately applied to new or replacement boilers; it is not applicable to the modification of existing boilers. It must be established that there is a sufficient market for new boilers to justify the effort and expense of developing a new boiler technology with improved pollution abatement potential.

Boiler systems have been built, operated, or proposed which incorporate fluidized bed combustion or gasification. Fluidized bed boiler concepts go back to 1928 when Stratton developed a spouting fluidized bed boiler (See Appendix C). The early boiler concepts incorporating fluidized bed combustion were generally developed for burning low-grade fuels. These designs did not recover heat or consider sulfur removal in the fluidized bed combustor. In the last decade fluidized bed boiler designs have been conceived, in England and in the United States, which incorporate heat recovery and sulfur removal in the fluidized bed combustor. Some of the concepts which have been proposed are reviewed in Appendix C.

Several surveys have been conducted to provide the information necessary to boiler design specifications, performance evaluations, and impact determination. Results from these surveys are summarized in Volume I. The survey data are presented in Appendices A and B in Volume III. These surveys include:

- Market surveys for industrial boilers and for utility power plants. These surveys have provided both information on past boiler sales and projections of future sales. Information on boiler capacities and on supply pressure and temperature has also been obtained.

- Fuel survey. Historical data on fuel usage in both industrial and utility boilers have been gathered. Evidence concerning fuel availability and estimates of present and future fuel prices have been collected and examined.

- Survey of SO_2 removal systems for cleaning combustion products of conventional boilers. Several proposed systems have been evaluated with regard to effectiveness and economy. A conventional boiler with a limestone wet scrubber for removal of SO_2 and particulates was selected as a basis of comparison for the performance of the fluidized bed combustion boilers designed in this work.

- Emission standards. Government standards are being established for limiting SO_2, NO_x, and particulate emissions from boilers. Projected standards have been used in setting functional specifications for the fluidized boiler designs.

Results from the various studies are utilized throughout this report.

FLUIDIZED BED COMBUSTION DATA

Fluidized bed combustion data on sulfur dioxide removal, limestone and dolomite regeneration, nitrogen oxide minimization, combustion efficiency, heat transfer, particle carry-over, boiler tube corrosion/erosion, and gas turbine blade corrosion/erosion were evaluated in order to establish operating and design specifications.

Sulfur Dioxide Removal

Sufficient limestone must be added to a fluidized bed combustor to reduce its SO_2 emission below that specified by the air pollution control laws in effect. The SO_2 emission level required to achieve pollution control objectives was discussed in the Preliminary Surveys section of the Summary Report. This section examines the relationship between the design selected for the fluidized bed combustor and the reduction in SO_2 emission which is likely to be achieved. This information is pertinent to the selection of operating conditions, since it is unlikely that control objectives will be selected independent of cost factors.

Work to determine the SO_2 removal achieved in fluidized bed combustors at various operating conditions has been carried out by a number of organizations, including Argonne National Laboratories (ANL); Esso Research and Engineering (Esso R & E); Pope, Evans and Robbins (PER); the National Coal Board (England), (NCB); Consolidation Coal Company (Consol); the U. S. Bureau of Mines; and Esso (England). (See [26,49,51,69,70,71, and 84] for details of the results obtained by each of these investigators to date.) No effort has been made in this study to tabulate all of the data acquired by these investigators with regard to SO_2 removal. Instead, an attempt has been made to:

- Identify the important factors affecting the degree of SO_2 removal
- Determine, at least semiquantitatively, the effect each of these factors has on the degree of SO_2 removal, recognizing that the effect of a change in any particular variable will be a function of the levels of many of the other variables

- Outline methods of extrapolating existing data to predict the SO_2 removal which might be achieved at operating conditions of commercial interest, including higher pressures, velocities, and temperatures.

Factors Affecting SO_2 Reduction

Thirteen variables which influence the percentage of SO_2 removal achieved in fluidized bed combustion of coal with limestone addition are listed in Table 1-1. Those variables at the top of the table are believed to exert stronger effects than those near the bottom. No claim is made, however, that the variables are listed in order of decreasing effect. The magnitude of the effect of each is a strong function of the level of the remaining twelve. Also shown in Table 1-1 are the ranges for each of the factors for which a significant amount of data has been reported and the approximate operating conditions for the industrial boiler and the atmospheric and pressurized utility boilers which are discussed in detail in later sections of this report.

One method used to summarize experimental data is the formulation of a mathematical model. A number of models for SO_2 removal by limestone and dolomite during fluidized bed combustion of coal have been proposed [49,70,84,100]. The model proposed by Koppel [49] can be used to illustrate the effects anticipated when some of the variables listed in Table 1-1 are changed. Specific mention will be made of those variables which are not clearly accounted for in this model.

The basic assumptions of Koppel's model [49] include:

- Plug flow of gas through bed
- Uniform SO_2 concentration in the direction transverse to the direction of the flow
- Perfect mixing of the solids within the bed
- Uniform temperature throughout the bed
- SO_2 generation throughout the bed with a generation function:

TABLE 1-1

Factors Affecting SO_2 Removal During Fluidized Bed Combustion

	Range for Which Significant Amount Of Data Are Available	Industrial Boiler	Atmospheric Utility Boiler	Pressurized Utility Boiler
1. Calcium-to-Sulfur Mole Ratio, Ca/S	1-4	6	6	6
2. Bed Temperature, °F	1400-1600	1650	1600	1750
3. Fluidizing Velocity, ft/sec	2-4	12.5	6-10	6-9
4. Stone Type & Source L = Limestone D = Dolomite	L-1359, L-T18, L - United Kingdom D-1337, D-Tymochtee D - United Kingdom	L-1359	L-1359	L-1359
5. Factors Affecting Gas-Solids Mixing (distributor design, placement of internals, fluidizing velocity/minimum fluidizing velocity, L/D ratio, location of stone feed)				
6. Factors Affecting Regenerated Stone Activity number of regeneration cycles, regeneration conditions, CO_2 partial pressure in combustor, amount of composition of ash				
7. Stone Size	25-1000 μ	- 1/4"	- 1/4"	- 1/4"
8. Bed Depth, ft.	1-4	2.5	2.5	10-16
9. Coal Type Swelling index, amount and composition of ash.	SI = 1 to 8	5-5.5	5-5.5	5-5.5
10. Location of Coal Feed and Means of Injection				
11. Combustion Pressure, atm.	1-5	1	1	10
12. Excess Air				
13. Sulfur Level of Coal, wt%	1-4.5	4	4	4

$$G(x) = \frac{\frac{FC_o}{L} e^{-\alpha x/L}}{1 - e^{-\alpha}} \qquad (1\text{-}1)$$

where

F = mass flow rate of gas through bed, lbs/min (includes CO_2, N_2, O_2, etc. after combustion);

C_o = lbs of SO_2 generated/lb of gas (after combustion) (assumed equal to the concentration of SO_2 at which $t_{0.63}$ was determined);

L = bed height;

x = distance above the distributor plate;

α = arbitrary parameter which describes where SO_2 is released ($\alpha = \infty$ corresponds to all SO_2 being generated at the base plate; $\alpha = -\infty$ corresponds to all generation at upper surface of bed; $\alpha = 0$, to uniform generation).

- The reaction at the particle is first order (irreversible) with respect to SO_2 concentration and at a fixed temperature is a function only of the degree of utilization of the stone, y, and particle size. The effect of particle size distribution is accounted for by using k(y) for the average bed particle size,

$$\frac{dy}{dt} = kC$$

$$k(y) = \frac{y_e}{C_o t_{0.63}} (1 - y/y_e) \qquad (1\text{-}2)$$

where

y_e = equilibrium loading of the sorbent, lbs SO_2/lb of CaO (not a function of gas composition);

$t_{0.63}$ = reaction time for a particle to reach 63% of its equilibrium loading ($C_o t_{0.63}$ is assumed independent of C_o).

The following implicit relationship was derived for R, the fraction of the total generated SO_2 that is removed by the sorbent:

$$R = \frac{1}{1 - e^{-\alpha}} \left[1 - \frac{\alpha \exp[-(1 - \frac{R}{r})H] - (1 - \frac{R}{r})He^{-\alpha}}{\alpha - (1 - \frac{R}{r})H}\right] \tag{1-3}$$

where

$r = \frac{wy_e}{FC_o}$, the effective ratio of additive feed to SO_2 feed;

w = feed rate of additive to bed, lbs of CaO/time;

and

$H = \frac{aLy_e}{FC_o t_{0.63}}$, the normalized bed height;

a = concentration of additive particles = $\frac{\text{lbs of CaO}}{\text{ft of bed height}}$.

Results generated from the above model formulation and previously published by ANL [49] indicate how both the fractional SO_2 reduction, R, and the particle consumption, C, can be expected to vary with the parameters H and r. C is related to stone utilization, U, by the equation:

$$C = U \frac{y_s}{y_e} . \tag{1-4}$$

Qualitatively, one anticipates the following effects based on the model:

1. Stone utilization decreases as the degree of SO_2 removal increases.

2. At values of r less than 0.5, the effect of Ca/S completely overshadows all other effects. (If $y_e/y_s = 0.5$, r = 0.5 corresponds to Ca/S = 1).

3. At values of H greater than 100, the degree of removal is also essentially determined by the value of r.

4. At low values of H (which might be caused by using shallow beds, beds dilute in CaO content, unreactive stone, or high gas velocities), the point of sulfur release determines the maximum degree of removal achievable. At high H values, the sulfur release pattern (unless $\alpha \to -\infty$) is unimportant.

5. The equilibrium loading achievable, y_e, may be the single most important characteristic of a given sorption system. This parameter enters into the calculation of both H and r. Its effect on H is especially important when H is low (see 4 above), and its effect on r appears important for all systems studied to date. Note in the following matrix that many of the operating variables (stone type, source, structure, and particle size, ash properties, and T) influence y_e. Difficulty in distinguishing between these effects can be anticipated.

6. Reaction kinetics for the stone-SO_2 reaction (quantified by the parameter $C_o t_{0.63}$) play a dominant role in determining degree of removal only if H is low (i.e., little gas-solids contacting time) or SO_2 is released near the top of the bed.

It is possible to estimate the change in SO_2 reduction caused by a change in any of the operating variables (except combustion pressure, excess air, and gas-solids contacting effects) if values of $t_{0.63}$ and y_e are known. The variations in $t_{0.63}$ and y_e caused by changes in temperature, stone type and source, particle size, and processing factors affecting stone structure must be correlated, and provision must be made for the unincluded factors in order to arrive at a complete picture based on Koppel's model.

EFFECT OF OPERATING CONDITIONS ON MODEL PARAMETERS

Factor	Model Parameters Affected
Ca/S	$r = Ca/S \times y_e/y_s$ where y_s = Stoichiometric stone loading
Temperature	Accounted for by changes in $t_{0.63}$ and y_e that occur with temperature
Fluidizing velocity	F, α, V for fixed cross-sectional area F/a, α, V for fixed bed composition and variable at fixed r bed area increasing V will decrease H
Stone type and source	$t_{0.63}$ and y_e
Factors affecting gas-solids contacting and solids mixing	Not included (See assumptions list 3 in text)
Factors affecting stone structure	Affect $t_{0.63}$ and y_e
Stone size	Affect $t_{0.63}$ and y_e
Bed depth	L, α
Coal type	α, also ash properties affect $t_{0.63}$ and y_e
Location of coal feed	α
Combustion pressure	May affect $t_{0.63}$ and y_e (However, this would be inconsistent with other model assumptions)
Excess air	Not included
Sulfur level in coal	No effect if all other variables maintained constant (This one is tricky, but remember that Co $t_{0.63}$ is constant by the assumptions of the model)

Summary of Reported Data

Calcium-Sulfur Ratio. Increasing the mole ratio of calcium to sulfur in the feed to the combustor, the Ca/S ratio, increases the degree of sulfur removal achieved.

There are many points in the system where calcium and sulfur might be measured, as is indicated in Figure 1-1. In the experimental investigations reported, Ca/S has most often been defined as the calcium content of the fresh stone feed (Stream A), in moles per unit time, divided by the sulfur content of the coal in Stream B, also in moles per unit time. The small amount of Ca which enters as coal ash is usually ignored because it is small in comparison with that added in Stream A. In this study Ca/S has been thought of as the mole ratio which includes calcium from A and C to the boiler. As such, it includes both Ca and S from the recycle as well as from the fresh feed. The two definitions are essentially identical when no regeneration is employed. The sulfur to calcium mole ratio at Point D is referred to as the degree of stone utilization, U.

The effect of Ca/S on the percentage of sulfur originally contained in the coal which is retained in the fluidized bed is illustrated in Figure 1-2. Percent retention is a nearly linear function of Ca/S at Ca/S ratios below 1.0. The Ca/S ratio required to achieve 90% sulfur removal is seen in Figure 1-3 to vary from approximately 2.3 for Welbeck Coal to greater than 3 for Humphrey Coal, all other conditions being similar.

Bed Temperature. Maximum sulfur removal has been obtained at bed temperatures between 1450°F and 1600°F for most coals, limestones, and operating conditions (other than temperature) tested thus far. Data on the effect of temperature on SO_2 removal reported by the National Coal Board, Argonne, and Esso (England) are plotted in Figure 1-3. Both the magnitude of the temperature effect and the absolute level of SO_2 removal achieved varied from investigator to investigator as a result of differences

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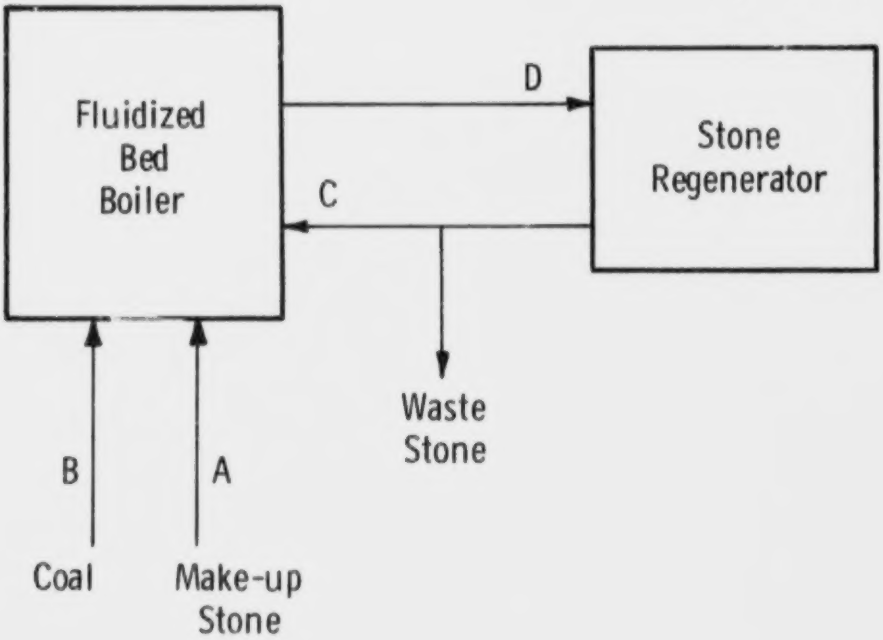

Fig. 1-1—Points on the system where calcium and sulfur levels can be measured

Curve 644303-A

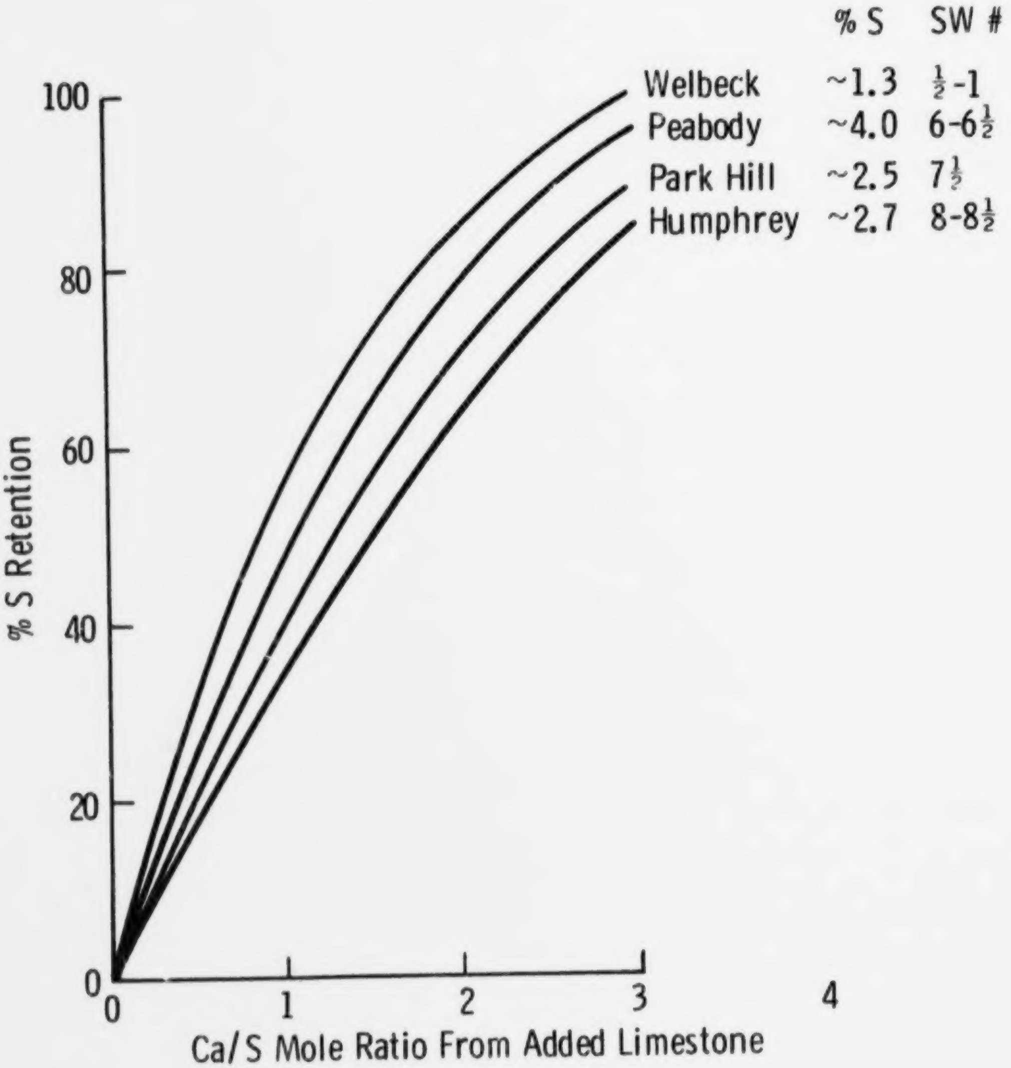

Fig. 1-2—Effect of Ca/S on sulfur retention for various coals. [Data from National Coal Board 69, 70 or 71]

in fluidizing velocity, fuel characteristics, stone characteristics, reactor design, etc. The data of all these investigators, however, show one common feature -- that a maximum in removal efficiency was observed rather than a monotonic increase or decrease in efficiency with temperature.

The presence of a maximum in the 1450°F to 1600°F range is not explained by available equilibrium and kinetic data on the primary reaction of concern:

$$CaO + SO_2 + 1/2\ O_2 \rightleftarrows CaSO_4. \qquad (1\text{-}5)$$

The equilibrium partial pressures for SO_2 range from approximately 10^{-8} atm at 800°C to 6×10^{-7} atm at 900°C (for 10% excess air). These values correspond (at 1 atm total pressure) to 0.01 and 0.6 ppm respectively. Since the actual exit gas compositions are several orders of magnitude higher (300 to 1500 ppm for the data points shown in Figure 1-3), the increase in equilibrium partial pressure with temperature does not explain the drop-off in SO_2 removal at high temperatures.

Kinetic data collected and analyzed by Skopp and co-workers [84] as well as by Coutant and co-workers [19] indicate that the rate constant (for reaction 1-5 preceded in some cases by calcination of $CaCO_3$) increases monotonically with temperature with a reasonable straight line fit to an Ahrennius plot.(1) Skopp investigated temperatures from 1200°F to 1600°F; Coutant's data were taken between 1500°F and 2150°F.

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(1) Once again caution should be taken before reaching sweeping conclusions. Harrington, Borgwardt, and Potter [40] characterized the reactivity of ten limestones and found that reactivity increased between 1600°F and 1800°F for four of them, but decreased for the other six.

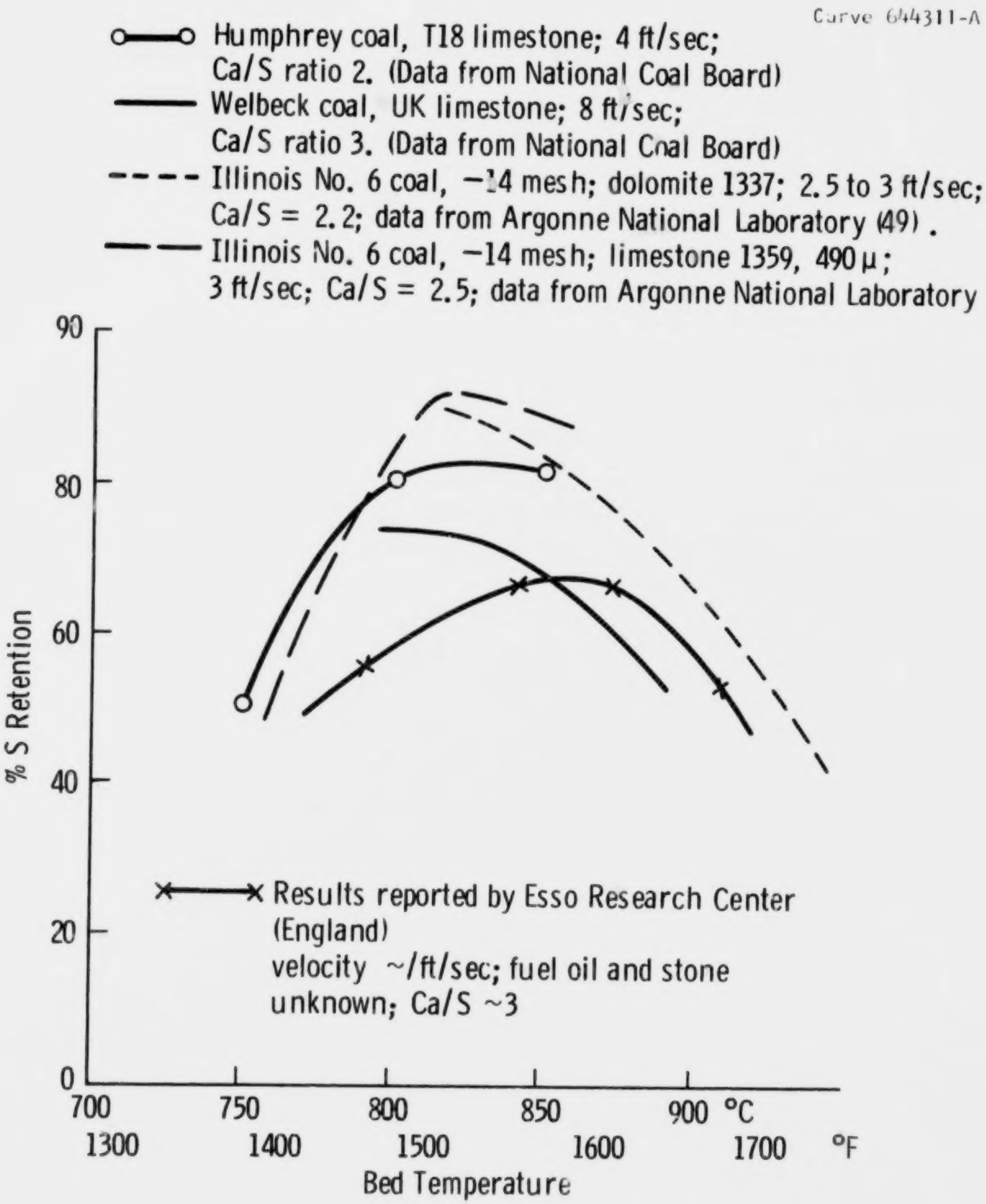

Fig. 1-3—Effect of bed temperature on sulfur removal in fluidized bed combustor

Two explanations for the reduction in SO_2 removal effectiveness at temperatures above 1600°F (Argonne [50] reported only 17% reduction at 1800°F) merit further consideration:

- An interaction between the coal ash and the stone which causes the formation of an inactive surface layer or the closing of the pores in the stone
- The presence of a localized reducing atmosphere in the emulsion phase of the lower portion of the fluidized bed combustor which results in a release of SO_2 from already sulfated stone.

Information to be considered in assessing the first explanation includes the following:

- Early work at Esso R & E indicated that "coal fly-ash does not stick or agglomerate with lime particles in the desulfurization reactor at 1600°F"[84, p.38].
- Later work at Esso R & E [38,39] indicated that the presence of coal ash during regeneration (∿ 2000°F)caused a dramatic reduction in the effectiveness of the regenerated stone for SO_2 removal. The decrease was attributed to "interference by components of the fly ash, possibly Fe_2O_3...."

Unfortunately, the above results shed no light on the importance of ash-stone interactions at 1700°F to 1800°F.

The second explanation has been proposed by workers at Argonne National Laboratories [50] and at Pope, Evans, and Robbins [80] and supported by some interesting experiments utilizing limestone to remove SO_2 during the combustion of natural gas. The theory, as expressed by Argonne, is:

> In any fluidized bed, an emulsion phase and a gas bubble phase are present. Gas in excess of that required for minimum fluidization is in the bubble phase but also circulates between the bubble and the emulsion phase. The gas in the bubble phase is essentially not available for reaction with solid particles until it moves into the emulsion phase. In the lower portion of a fluid-bed coal combustor, oxygen in the emulsion phase reacts very rapidly with coal. All of the oxygen in the emulsion phase is consumed, forming carbon monoxide. As the bubbles rise through the bed, air exchanges between the bubbles and the emulsion phase; in the upper part of the bed, the emulsion phase contains excess oxygen.
>
> Thus, in the lower part of the combustor, the following reaction might occur:
>
> $$CaSO_4 + 4\ CO \rightleftarrows CaS + 4\ CO_2 \quad . \tag{1}$$
>
> In the upper part of the combustor, one or more of the following reactions might occur:
>
> $$CaS + 3/2\ O_2 \rightleftarrows CaO + SO_2 \tag{2}$$
>
> $$CaS + 3\ CaSO_4 \rightleftarrows 4\ CaO + 4\ SO_2 \tag{3}$$
>
> $$CaS + 2\ O_2 \rightleftarrows CaSO_4 \ . \tag{4}$$
>
> Reactions 2 and 3 would result in SO_2 being regenerated in the upper part of the combustor, seriously limiting the efficiency of the system for retention of sulfur. These SO_2 regeneration reactions are highly temperature dependent and might account for sulfur retention efficiency being lower at higher temperatures.

While the reaction sequence proposed above is certainly speculative, the theory that SO_2 is released as a result of localized reducing conditions in the lower portion of the fluidized bed is supportable. Investigators at Pope, Evans, and Robbins have demonstrated that $CaSO_4$ can be decomposed to CaO and SO_2 at 1800°F to 1900°F, even when the overall feed rates of fuel and air indicate an oxidizing atmosphere exists [26], and have proposed an in-bed regeneration concept based on this information. Argonne [50] acquired data which showed that essentially no SO_2 was liberated from $CaSO_4$ under slightly oxidizing conditions at 1625°F, but an appreciable amount was liberated at 1720°F and above. At 1850°F, the partial pressure of the SO_2 liberated was a strong function of the O_2 partial pressure in the flue gas, varying from 2700 ppm at 1.8% oxygen to 4900 ppm at 0.5% oxygen.

Summarizing, the poor SO_2 removal efficiency obtained at temperatures near 1800°F seems due primarily to the liberation of SO_2 as a result of local reducing conditions in the bottom section of the bed. Stone-ash interactions may also contribute to the loss of effectiveness, especially in the case of high-ash coals [96]. If stone-ash effects do not dominate, the use of deeper beds and higher excess air is expected to result in better removal at high temperatures.

The experimental results of Zielke, et al.[106] can be cited in discussing the effect of bed depth at high temperatures. They reported SO_2 reductions ranging from 75% to 97% at 1800°F, in sharp contrast with all others who have reported high temperature results. Their tests were carried out in a 4-inch diameter bed with a depth of 36 in, an L/D of 9. In contrast, Argonne National Laboratories operated with an L/D of 4 and Pope, Evans, and Robbins with an L/D of approximately 1. The authors [106] attributed their good results to good solids-gas contacting, but Ehrlich [26] pointed out that the high L/D might have reduced the degree of solids mixing and thereby have produced the encouraging results. According to the latter theory, localized reducing conditions exist in the lower section of the deep bed, but the gas composition has evened out in the top section so that the dolomite there is effective. It is impossible to choose between the above explanations on the basis of current data. In fact, it seems likely that both factors contributed to the improved performance.

Fluidizing Velocity. If all other factors remain unchanged, increasing the fluidizing velocity decreases the effectiveness of sulfur removal. Typical results are shown in Figure 1-4. Although increasing the Ca/S ratio diminishes the effect, even at Ca/S = 4 the experimental results indicate a decrease from 90% removal efficiency at 2 ft/sec to 70% at 6 ft/sec. It should be noted that all Ca/S ratios shown are for fresh stone and that higher Ca/S ratios will be required to achieve the same effect with regenerated stone.

Increasing the gas residence time in the fluidized bed by increasing bed height can be anticipated to compensate for increased velocity, provided that gas-solids contacting remains effective.

Stone Type and Source. Both limestone and dolomite have been utilized as SO_2 sorbents in fluidized bed combustion systems. The distinction between the two, for purposes of this report, is in their $CaCO_3$ content: a limestone contains more than 80% $CaCO_3$ while the dolomites investigated have had nearly equal molal amounts of $CaCO_3$ and $MgCO_3$. At a fixed value of Ca/S, dolomites have been found typically to achieve a higher sulfur removal; since only about half of the dolomite consists of Ca compounds, however, dolomites typically achieve a lower degree of sulfur removal than limestones at the same value (lbs of stone/lb of S).

A more important and more fundamental difference between the two stone types is based upon the phase diagram shown in Figure 1-5. MgO is the stable Mg-compound at almost all temperatures and operating pressures being considered for fluidized bed combustion. CaO is stable only at the conditions in the lower right-hand corner of the plot -- that is, at relatively high temperatures and low pressures. At a total combustion pressure of 10 atm and 1600°F, for example, limestone fed

Curve 644307-A

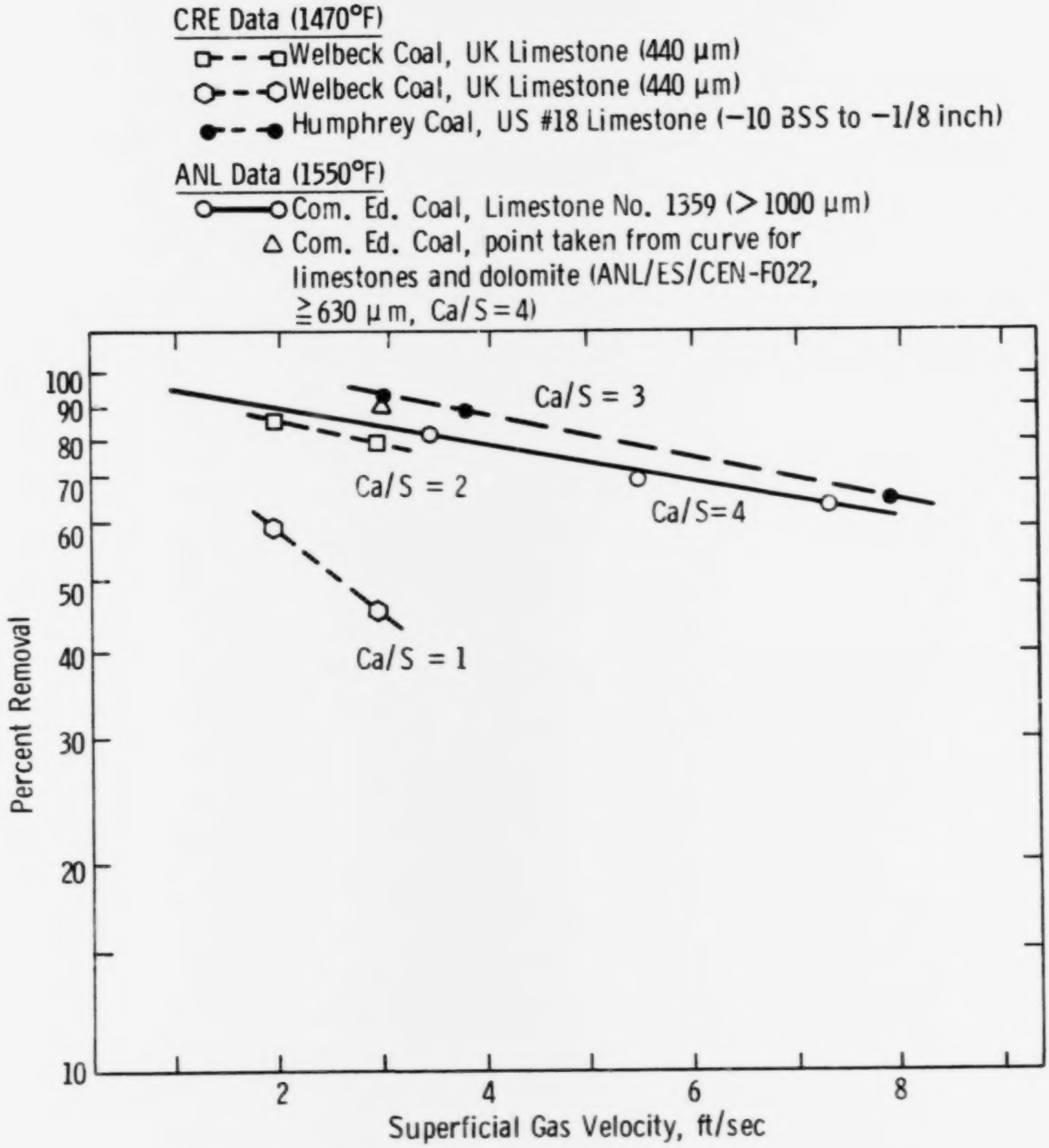

Fig.1-4—Effect of superficial gas velocity on SO_2 removal

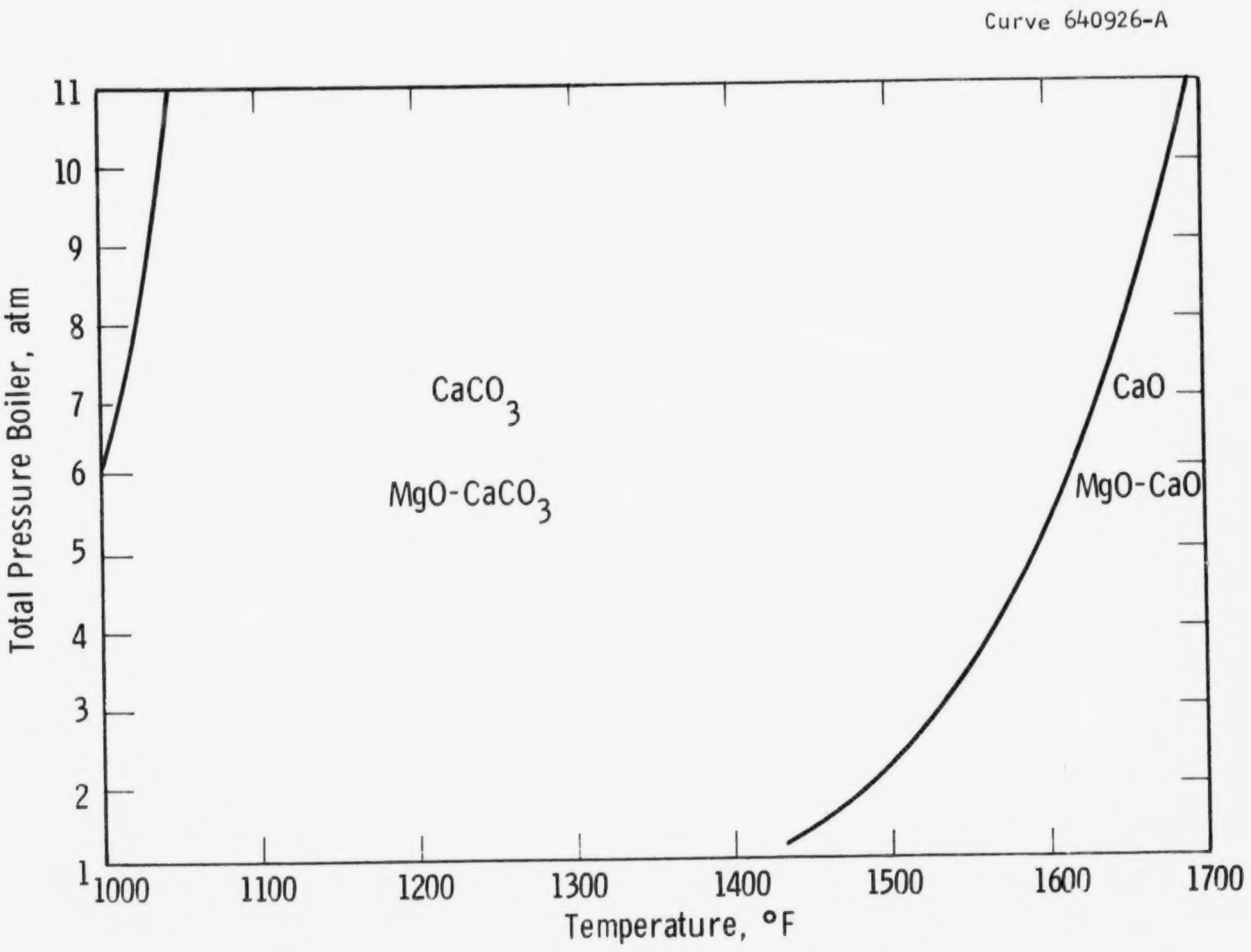

Fig. 1-5-Phase diagram for limestone and dolomite, assuming a boiler atmosphere containing 15% CO_2

to the combustor would not calcine and, therefore, would not be an effective SO_2 sorbent. At the same temperature and pressure, dolomite would be readily converted to the "half-calcined" state: $MgO-CaCO_3$. This provides the necessary stone porosity for effective SO_2 sorption in high-pressure boilers, unless the operating temperature (even during turn-down) would be maintained at 1680°F or higher.

Although there is no evidence that MgO reacts with SO_2 in the fluidized bed, its presence appears to account for the better utilization of CaO in dolomite (as compared to limestone) which was noted previously. Data obtained by Skopp and co-workers [84] and shown in Figure 1-6 illustrates this point. The reaction rate constant at a given utilization increased monotonically as the ratio MgO/CaO was increased in these experiments. Apparently, the presence of unreacted MgO helps to maintain an open structure so that more of the Ca can be reacted.

This conclusion is supported by analysis of partially reacted limestone and dolomite by Jonke and co-workers [49] and by the National Coal Board [71]. NCB utilized mercury porosimetry to examine the porosity of the particles and reported that an impermeable sulfate shell was formed around limestone 1359 particles but not around particles of dolomite (either that from the United Kingdom or U. S. No. 1337). Jonke and co-workers at ANL confirmed this result when SO_2 and air were injected into a fluidized bed of limestone. When SO_2 was generated by combustion of coal in the fluidized bed, however, no shell was observed around the limestone particles. (ANL utilized electron microprobe analysis in their work.) This difference between results obtained with SO_2 generated by combustion and SO_2 injected with air was attributed to the limestone's cycling between oxidizing (absorbing) and reducing (desorbing) conditions as it circulated through the combustion bed. The alternate adsorption-desorption, it was postulated, permits the SO_2 to penetrate throughout the particle. (It also reduces overall percent SO_2 reduction achieved in shallow beds.) The above result may explain the fact that no statistically significant difference in SO_2 reduction (at $Ca/S \simeq 2.3$)

Curve 644306-A

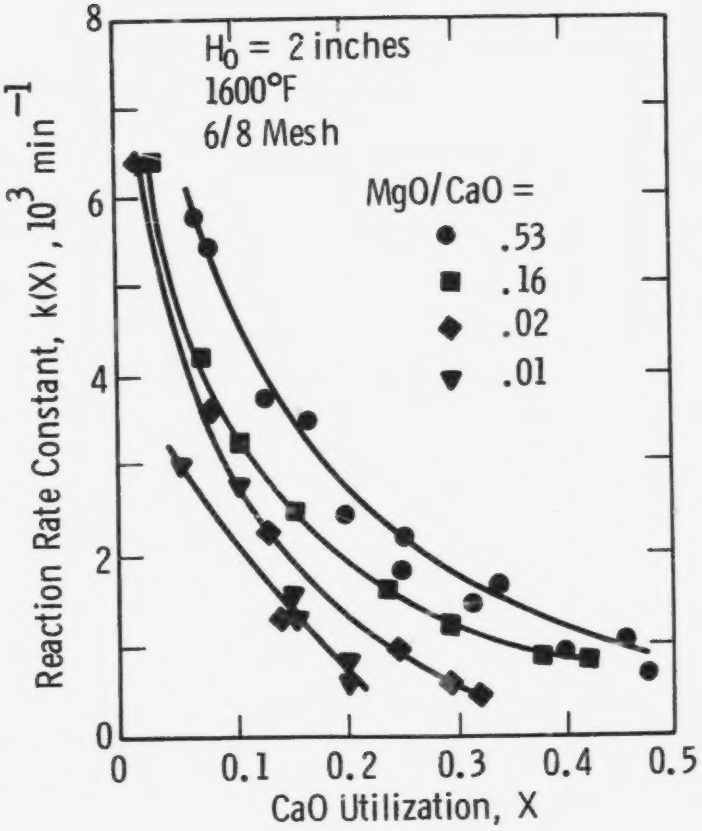

Fig. 1-6—Effect of MgO/CaO ratio on stone reactivity

achieved with dolomite 1337 and limestone 1359 was observed in the ANL 6-inch diameter fluidized bed combustor [51], even though significant differences in reaction rate constants have been observed [49] on synthetic flue gases.

In addition to the differences in SO_2 removal efficiency obtained with limestone versus dolomite, stones of the same type but from different sources show wide variations in their reactivity with SO_2 [19,41,63,71,80,84]. In fluidized beds operating at 1450°F to 1600°F, it has been found that U. S. Limestone No. 18 has a higher reactivity than U. K. Limestone, which has a higher reactivity than U. S. Limestone No. 1359 [71]. Tymochtee dolomite has been found more effective than U. S. Dolomite 1337, especially at temperatures higher than 1600°F [49,51]. As illustrated in Figure 1-7, the magnitude of the differences between stones is large. These differences may be due to differences in the rates of calcination, differences in the pore size distribution present after calcining, subtle differences in chemistry, interactions between the coal ash and the stone (see Figure 1-8) or to a combination of these and/or other effects. Numerous studies have been made [10,19,70,71,80] and are in progress to elucidate the reasons, but no clear picture has emerged at this time.

Another important difference between stones is their resistance to attrition. This is especially important when a regenerable system is considered. The best results reported to date have been by Consolidation Coal Company for Tymochtee dolomite [106]: 0.25% to 1.5% calcium attrition per regeneration cycle. Other results include 3% to 5% attrition per cycle for limestone 1359 [84] and 10% per cycle for U. K. Limestone.(1)

In view of its high sulfur retention, better regeneration characteristics, and low attrition race, Tymochtee dolomite appears to be an attractive sorbent and should be studied further in any future large-scale fluidized bed combustion experimentation. The success of future

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(1) Esso Research Centre, England, private communication.

Curve 644304-A

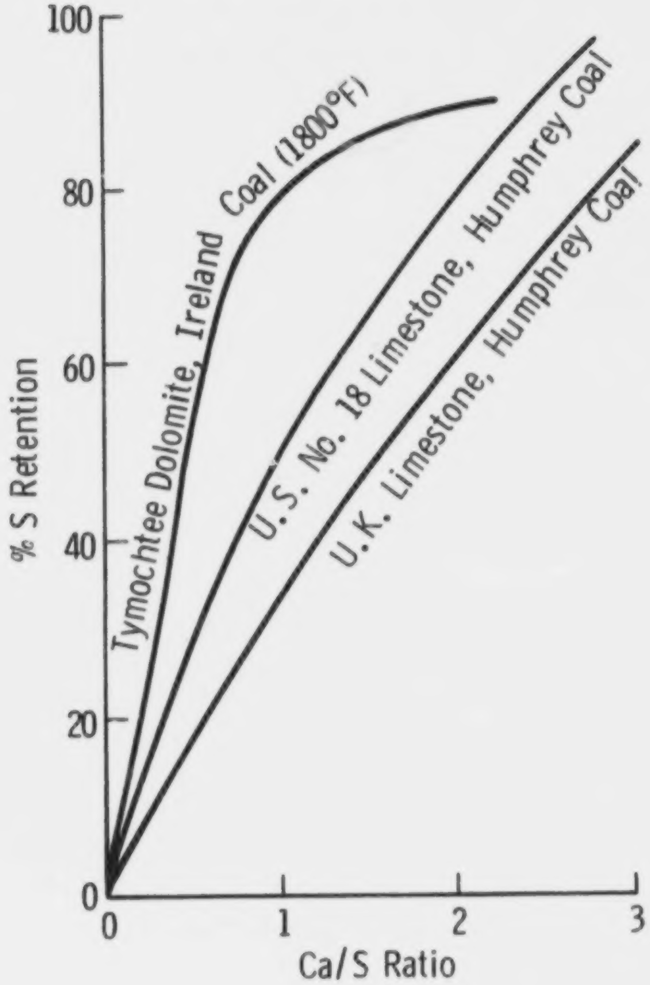

Fig. 1-7- Sulfur removal effectiveness

Curve 644305-A

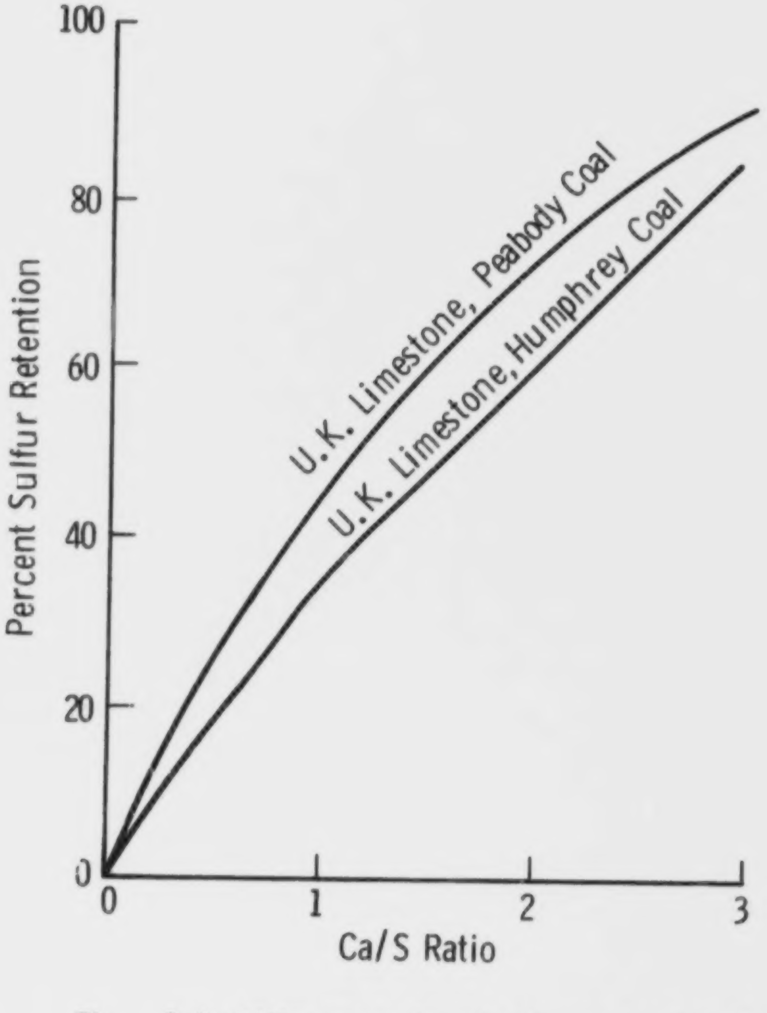

Fig. 1-8—Differences in effectiveness of sulfur removal with same limestone, but different coals [Data from National Coal Board 69]

fluidized bed combustors in effectively reducing SO_2 pollution may depend upon finding stones with properties similar to Tymochtee dolomite near the proposed power plant site.

Regenerated Sorbent Activity. Available data indicate that sulfated limestone and dolomite can be regenerated at atmospheric pressure for subsequent sulfur absorption duty. Calcium sulfate is reduced to calcium oxide while driving off an SO_2-enriched gas stream -- 3% to 10% SO_2. No data are available for sorbent activity following the two-stage regeneration forming $CaCO_3$ plus H_2S for regeneration at high pressures.

PER performed simultaneous absorption-regeneration runs in their fluid bed module equipped with carbon burn-up cell (CBC) (Monthly Progress Report 18, March 1971). Although the CBC would not accomplish simultaneous carbon burn-up and sorbent regeneration (PER concluded that three reactors are needed: a boiler, a CBC, and a regenerator), stone activity remained high. One bed operated in excess of 100 hours achieved 90% sulfur removal. Limestone addition was needed to replace analytical samples and attrition losses -- approximately 5% of the coal rate. This method of operation corresponds to that expected in an actual boiler system. Sorbent is continuously fed to and withdrawn from the boiler bed, regenerated and recycled back. All beds contain some fresh stone and some stone that has undergone many absorption-regeneration cycles. Regenerator operating conditions, however, will have to be modified, since the SO_2 level in the regenerator off-gas was usually below 3%.

Esso R&E [84] examined the cyclic capacity of 1359 limestone and 1337 dolomite in batch absorption-regeneration experiments. In the first set of experiments a bed of solid sorbent was sulfated by a simulated flue gas. Simulated regeneration gas converted $CaSO_4$ back to CaO, driving off SO_2. The results from these experiments are shown in Figure 1-9 as curves 1 and 2. The limestone activity increased after the first regeneration cycle and then steadily decreased. Dolomite activity measurements were difficult to obtain because of high attrition losses.

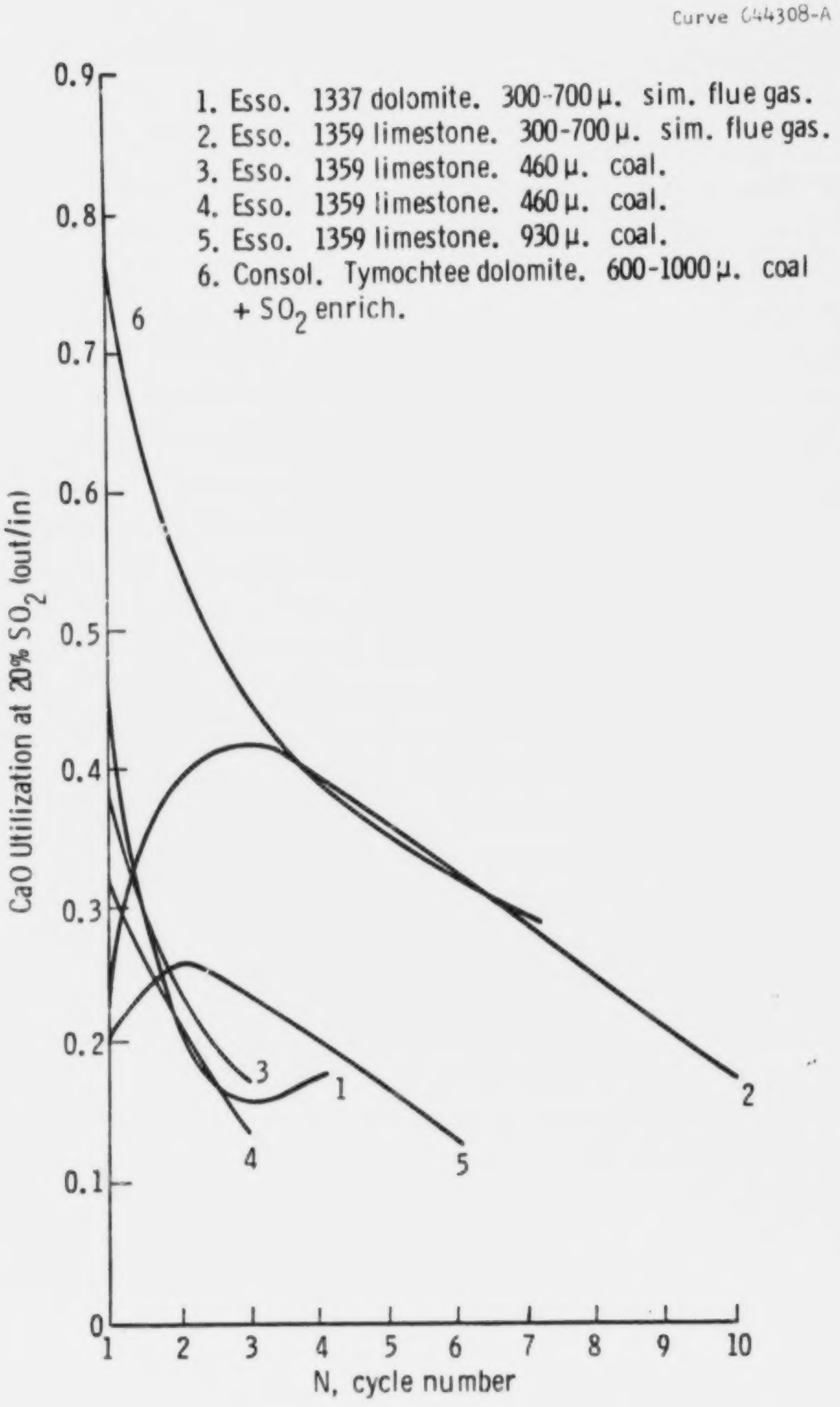

Fig. 1-9—Regenerated sorbent activity

In Esso's second set of experiments, a bed of 1359 limestone absorbed sulfur released from combusting coal in the bed. After sulfur absorption is complete, the bed is regenerated by reacting with simulated regeneration gas. The bed is then sent back to the combustion chamber for another sulfur absorption cycle. Sorbent activity for these runs is much lower than when simulated flue gas was employed (see Figure 1-9, curves 3, 4, and 5). An interaction between the Fe of the coal ash and the stone was postulated as the cause of reduced activity.

Zielke and co-workers [106] examined the cyclic activity of Tymochtee dolomite. Their procedure was somewhat different from Esso's in that they maintained a continuous sorbent flow through the combustor and regenerator beds. Combustion and regeneration were not done simultaneously, but rather in batches. In absorption runs the fluidizing gas was enriched with SO_2 to achieve the desired sulfur absorption rate -- 80%. Thus, the SO_2 level in the absorption bed is much higher than in the Esso or PER experiments. Zielke's results are shown in Figure 1-9 as curve 6.

For a conservative design basis, one could select the CaO utilization following regeneration bounded by curves 1, 3, and 4. This would suggest that the sorbent could undergo five to six regeneration cycles and still retain sulfur absorption activity. Curves 2 and 6 seem to give an upper limit on the CaO utilization that might be achieved following regeneration.

The regeneration processes have to be studied in more detail to determine the effects of regeneration on the sulfur absorption properties of the sorbent. Conceptually, various regeneration conditions may alter the size, structure, chemical composition, surface properties and strength of stone entering the boiler. At this point in time we can only speculate on which regeneration factors are significant. The next matrix gives a preliminary listing of regeneration conditions we believe will have a significant effect on sorbent properties.

REGENERATOR FACTORS

Regenerator temperature	Average and maximum to which sorbent is exposed
Regenerator pressure	Heat transfer to sorbent and partial pressure of gaseous components
Gas composition	Oxidizing or reducing; calcining or carbonating; steam-hydroxide melt formation; local and overall
Solids residence time	Exposure time to high temperature; degree of regeneration
Gas velocity and bed depth	Attrition
Regenerator fuel	Ash level in bed; carbon coating on particles

Stone Particle Size. Stone particle sizes utilized in removing SO_2 during fluidized bed combustion can be divided into two classes:

- Coarse -- stone particles large enough so that they are not elutriated and thus make up the dominant material in the fluidized bed. Such stones generally have a mean particle diameter greater than 300 μ.
- Fine -- stone is completely elutriated from bed; bed consists primarily of an inert material. Such stones generally have a mean particle size less than 100 μ.

There is no lack of disagreement in the data reported:

- In shallow beds (10-18 in) at high velocity (13 ft/sec), PER [80] reported that fine stone gives higher SO_2 removal than does coarse stone. (See Figure 1-10)
- In deeper beds (3 ft) at low velocity (3 ft/sec) the NCB [70,71] reported higher removal with coarse stone than with fine. (See Figure 1-11)
- To cover all combinations, ANL [44] reported comparable results with coarse and fine stone.

Closer examination shows that the fine particle data from all experimenters is in agreement, despite the differences in velocity. The low SO_2 reduction obtained by PER is believed to be a direct result of their experimental technique and equipment geometry. First, their

Curve 645931-A

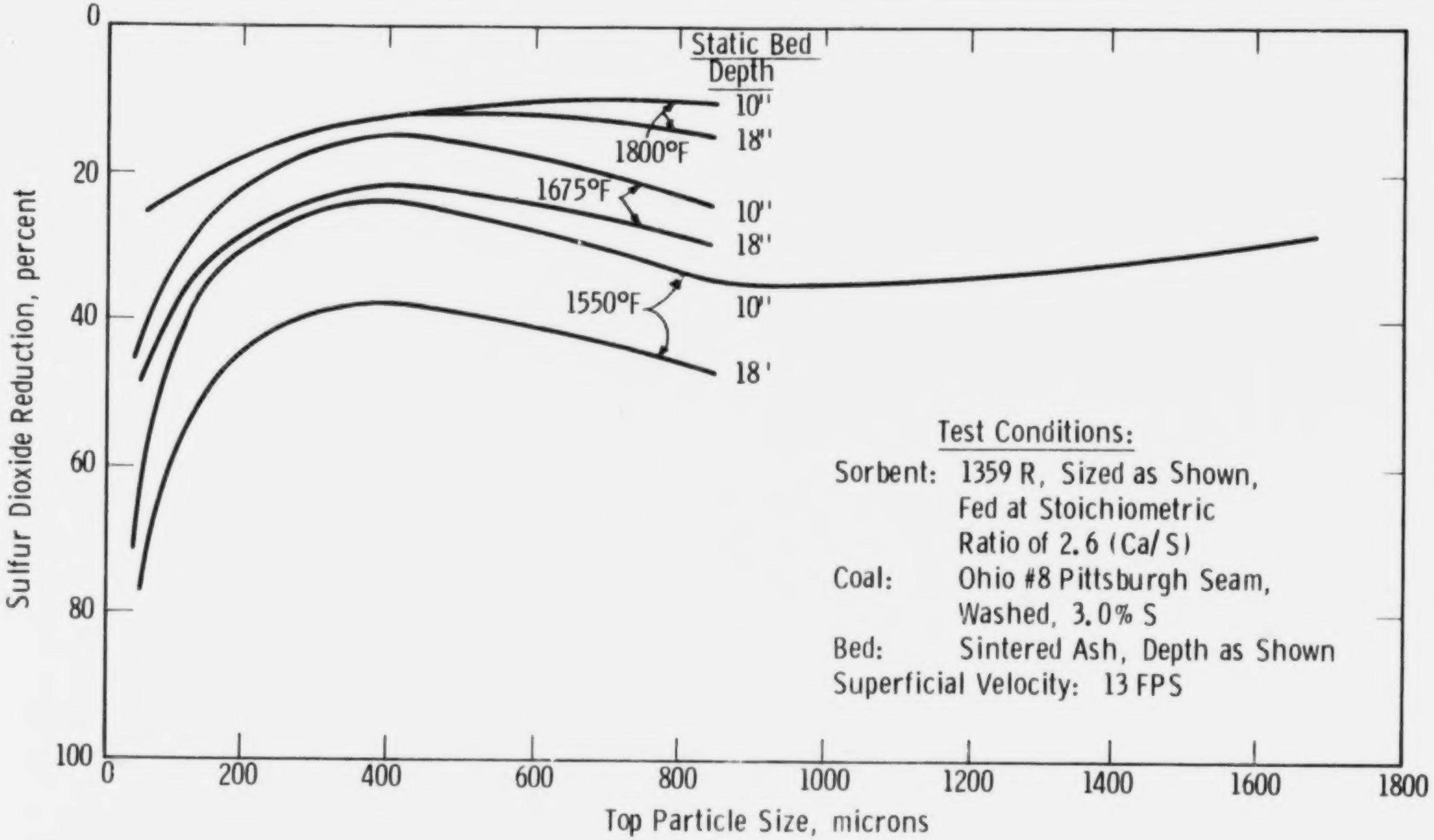

Fig. 1-10—Variation in sulfur dioxide reduction with sorbent particle size, bed depth and bed temperature [80]

Curve 645839-A

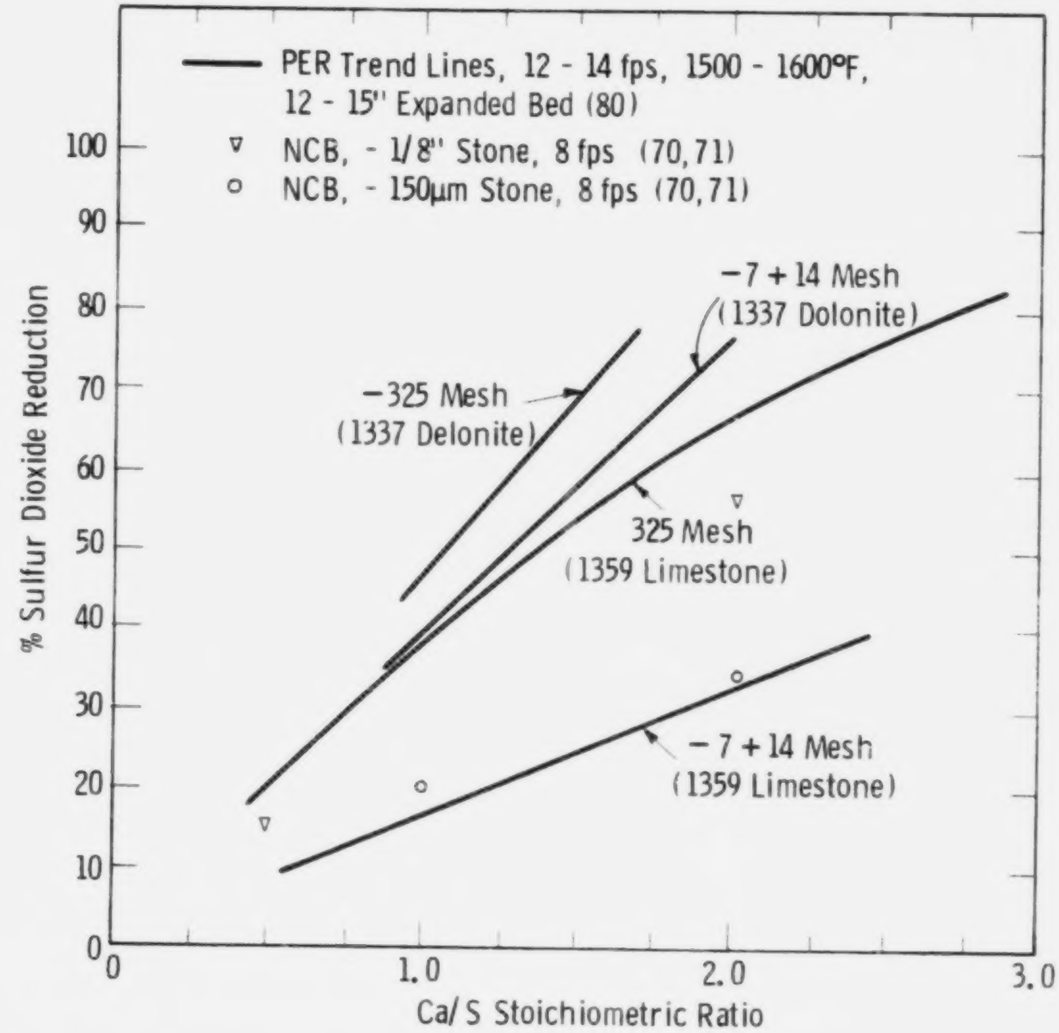

Fig. 1-11—Effect of stone size on sulfur dioxide reduction

experiments were started with an ash bed and were not run long enough to achieve a predominantly stone bed. Thus, referring to Equation 1-3, the product 'aL' in the definition of normalized bed height was low. Also, because of the extremely low aspect ratio employed, much of the SO_2 would be expected to be released near the top of the bed in their apparatus.

Coarse particles are recommended primarily because regenerable systems seem necessary. The ash from most American coals investigated is elutriated from the bed. Therefore, if coarse stone which remains in the fluidized bed is utilized, separation of the stone and ash can be achieved in the fluidized bed combustor, thus simplifying the regeneration process. In addition, if the results of PER are discounted for the reasons noted above, it appears that the increased residence time of the coarse stone in the system more than compensates for its reduced reaction rate constant so that the SO_2 removal achieved at a given Ca/S can be expected to be as high (or higher) with coarse stone as with fine (except where low L/D and high temperatures are employed).

Bed Depth. Altering the bed depth in a combustor of fixed cross-sectional area changes the gas residence time in the bed, solids circulation, and the aspect ratio (L/D) for the bed. Reported data are of little value in predicting the effect of bed height on SO_2 removal in contemplated boiler designs. It is useful, however, in illustrating that both residence time and aspect ratio effects are important.

Esso reported [38,39] that increasing settled bed height (from 4 in to 17 in) decreased the stone utilization fixed value of SO_2 removal (from 20% to 7%). Both their calculations and experimental observations indicated that the bed was slugging badly. In this case the effect of poorer gas-solids contacting completely overshadowed the apparent increase in residence time.

Pope, Evans, and Robbins [80] reported an increase in SO_2 removal from 35% to 47% at 1550°F and from 24% to 30% at 1675°F when the static bed height was raised from 10 in to 18 in (See Figure 1-10). These increases are approximately 70% of those predicted from a simplified model (i.e., $R = 1 - e^{-\gamma L}$). Once again, however, it must be pointed out that these experiments were started with an ash bed and were of short duration.

The inventory of stone in the bed (and thus the product 'aL') was, therefore, more a function of run time than bed height. Furthermore, aL cannot be calculated from the data published. The mechanistic reason for the improved removal remains undefined.

The small improvement observed in Figure 1-10 by increasing L with fine particle systems is not surprising, since most of the stone is carried along with the gas rather than maintained in the bed. The gas contact with stone is determined more by the total reactor size (including freeboard) than by the bed depth.

Bed Area. The cross-sectional area of the fluidized bed is also an important factor to consider in data evaluation. The bed areas for the proposed designs range from 35 ft^2 for each bed in a 300 MW pressurized boiler plant to over 300 ft^2 for each bed in a 600 MW atmospheric pressure plant. Data have been acquired on beds ranging from 0.1 ft^2 to 9 ft^2. Wall effects, solids circulation patterns, and gas flow characteristics change with bed diameter. Preliminary indications of the effect of scale-up are encouraging: essentially the same SO_2 removal was obtained in a 3 ft x 3 ft square bed as had been obtained in a 6-in diameter bed using the same materials at identical operating conditions.

Coal Type. The degree of SO_2 removal achieved may be influenced by many coal properties: swelling index, ash composition, percent ash, ratio of inorganic to organic sulfur, total volatiles content, and particle size, to name a few. Experimental evidence has been presented to support the effect of the first two properties.

Figure 1-2 illustrates the fact that better SO_2 removal is achieved with coals having low swelling indices than those having high ones. This can be explained by the faster release of volatiles from the coals with low swelling indices. Thus, α is higher for coals with low swelling indices. Argonne [49] has documented the real effect of sulfur release patterns on the removal. However, the importance of the sulfur liberation pattern from the coal itself would seem unimportant at bed temperatures exceeding 1600°F where sulfur is constantly being released and readsorbed by the adsorbent. (See previous section on the effect of temperature.)

The effect of coal ash components on the activity of the stone after regeneration was discussed in previous sections.

The Bureau of Mines [95,96] reported the formation of a glassy coating resulting from the reaction of silicate minerals in the coal ash with lime. The run was made at 1505°F using a high-ash (Lower Kittanning), unwashed coal. The sulfur removal obtained in this run with Ca/S = 1.4 was only 15% compared to results near 70% for low-ash coals in the same series of runs. In subsequent runs with washed Lower Kittanning coal, however, high sulfur removals (89% to 94%) were obtained. Further investigation of the interaction between limestone and silicate minerals seem justified, but no firm conclusions with respect to its effect on SO_2 removal can be drawn at this time.

Excess Air. The amount of excess air appears to play a minor role in determining SO_2 removal at lower (below 1500°F) combustor temperatures, but a major role at higher temperatures. This effect is believed to result from inhibiting the re-release of SO_2, which occurs when locally reducing conditions exist (see discussion of temperature earlier in this section). However, no quantitative data on the effect of high excess oxygen (10% to 50%) at elevated temperatures (1600° to 1900°F) were found. Such data may be vital to effective design of a carbon burn-up cell with low SO_2 emissions.

Other Factors. Other factors influencing SO_2 removal, but not discussed as separate topics in this subsection, include:

1. Location of coal and stone feed and their method of injection
2. Gas-solids contacting effectiveness
3. Extent of solids circulation within the bed
4. Combustion pressure
5. Sulfur level of the coal.

The effects of factors 1, 2, and 3 above have been alluded to in earlier parts of this subsection. Sufficient data are not available to quantify these effects on SO_2 removal, but a summary of the qualitative picture will be made here.

Coal Feed. The coal should be fed in such a way as to minimize the amount of SO_2 released near the top of the bed. This can be done by injecting it near the bottom of the bed and by preventing its being swept upward by the circulation patterns in the bed immediately after injection.

Stone Feed. The stone should be fed near the top of the bed to achieve maximum counter-current action.

Gas-Solids Contacting Effectiveness. Good contacting is achieved by preventing channelling and the formation of large gas bubbles, i.e., by maximizing the interchange of gas between the emulsion phase and the bubble phase of the bed. Accomplishing good contacting is one of the keys to designing fluidized beds and is the subject of an entire section of chemical engineering literature. (See, for example: Leva [61], Kunii and Levenspiel [60], Zenz and Othmar [105].)

Solids Circulation. This factor is likely to be much more important at high-temperature (i.e. > 1550°F) bed operation than at low-temperature operation. Since the combination of high temperature and locally reducing conditions results in the release of SO_2 from $CaSO_4$, it is desirable to reduce the probability of locally reducing conditions and also the amount of $CaSO_4$ present near the top of the bed when high temperatures are employed. This can be achieved by reducing the degree of solids mixing, if the fresh stone is fed near the top of the bed and coal near the bottom. Those factors which improve contacting effectiveness, unfortunately, also tend to improve solids circulation unless special baffles or other intervals are utilized. Once again, the method employed to achieve a balance is best left to the bed designer.

Combustion Pressure. If the reaction of CaO with SO_2 is first order in P_{SO_2}, then operating pressure will affect the degree of SO_2

removal achieved only via its effects on the calcination rate of the sorbent and on the quality of fluidization. British results [69, Fig. 7, Task II] in which greater than 90% reduction in SO_2 emission was obtained with an average Ca/S of approximately 2 and an operating pressure of 5 atm are encouraging. The low velocity (∿ 2 ft/sec) and temperature (1480°F) employed, however, and the limited data available make extrapolation of these results to the proposed conditions difficult.

Sulfur Level of the Coal. Once again, no effect is anticipated if the reaction is first order. Data reported (see Figure 1-2) have shown no correlation between sulfur level in the coal and the degree of removal achieved. One caution: at high operating temperatures where the amount of SO_2 released from $CaSO_4$ is high compared to the amount released by the coal, it seems likely that the degree of removal will be directly proportional to the sulfur level in the coal. In other words, the SO_2 level in the gas and, therefore, the emission will be determined completely by the temperature, gas composition, stone loading, etc.

Extrapolation of Existing Data

Despite the complexity of interaction among the effects of the variables which influence sulfur removal, it appears feasible to extrapolate a limited amount of existing data.

The general form of the Koppel model outlined previously is recommended for this use. Although it is relatively simple mathematically and contains only three experimental parameters (y_e, $t_{0.63}$, and α), this model form appears to predict correctly the direction and order of magnitude of change in SO_2 removal which result from changes in Ca/S, fluidizing velocity, bed depth, and coal type. It has the flexibility to project the in-bed effects of temperature, stone type and particle size, regeneration (and other factors affecting stone structure), pressure, and excess air on the basis of a limited amount of bench-scale experimental work. Slight modifications are suggested to account for contacting effectiveness, the re-release of SO_2 from $CaSO_4$ owing to locally reducing conditions, and the use of partially sulfated stone. Also suggested is a method for treating deep beds and beds with internals where solids circulation cannot be expected to be "perfect."

The degree of SO_2 removal projected by the model for the industrial boiler design operating conditions illustrates the approach. The industrial boiler design has the following operating conditions:

$T = 1650°F$

$L = 2.5$ ft

$v = 12.5$ ft/sec

$\overline{d_p} \approx 2000\ \mu$

limestone 1359

$P = 1$ atmosphere

Pittsburgh #8 seam coal

4.3% sulfur

Swelling index 5 to 5.5.

Argonne's data point at 1600°F for Illinois No. 6 coal, 490 μ 1359 Limestone, 3 ft/sec., Ca/S = 2.5, static bed height = 15 inches, and utilizing partially sulfated stone as the bed material (shown in Figure 1-3) was chosen. This data point was chosen because it employed the same sorbent, the same bed material, a similar coal, and nearly the same temperature. Also, this run had continuous feed and withdrawal of reactants, it was allowed to reach a steady operating condition, and its results are consistent with other well-defined runs at similar conditions.

A plot of rate constant versus CaO utilization can be used to determine (y_e/y_s). The intercept of the straight line with the utilization axis gives the value of y_e/y_s consistent with Koppel's expression for $k(y)$:

$$k(y) = \frac{y_e}{C_o t_{0.63}} (1 - y/y_e). \qquad (1\text{-}6)$$

Data for dolomite 1337 [41] are plotted in Figure 1-12. Similar plots for limestone 1359 kinetic data obtained by Esso [84] indicated values of y_e/y_s of 0.4 for -16/+18 mesh material (∿ 1000 μ) and $y_e/y_s = 0.25$ for -6/+8 mesh (∿ 2400 μ). All of the above data were taken at 1600°F. A plot of y_e/y_s versus average particle diameter is shown in Figure 1-13.

If one assumes that particles in a bed of burning coal have the same value for y_e as those tested in synthetic flue gases, then it is proper to utilize $y_e/y_s = 0.65$ for the 490 μ particles used in Argonne's experiment and 0.27 for the industrial boiler which employs stone with an average diameter near 2000 μ. If localized reducing conditions exist, however, it has been shown that sulfur is uniformly distributed throughout the particle, implying the possibility that $y_e/y_s \rightarrow 1$. As a compromise, a value of $y_e/y_s = 0.9$ was used for the Argonne experimental data and $y_e/y_s = 0.7$ for the commercial unit.

Equation 1-3 simplified for $\alpha = 0$ (i.e., uniform sulfur generation throughout the bed) was utilized in the calculations

$$R = \frac{(1 - R/r)H + e^{-(1 - R/r)H} - 1}{(1 - R/r)H} \qquad (1\text{-}7)$$

Substituting $R = 0.86$ and $r = 2.5 \times 0.9$ into Equation 1-7, H can be calculated for the Argonne run as $H_{exp} = 11.5$. If H were calculated directly from its definition and available kinetic data, a value of 58 is obtained.

The difference between H_{exp} and H can be explained in terms of gas-solids contacting effectiveness and the reduction in net SO_2 pickup due to locally reducing conditions, i.e.,

$$H_{exp} = H \times \varepsilon \times \frac{C_o - C^*}{C_o} \qquad (1\text{-}8)$$

Curve 644309-A

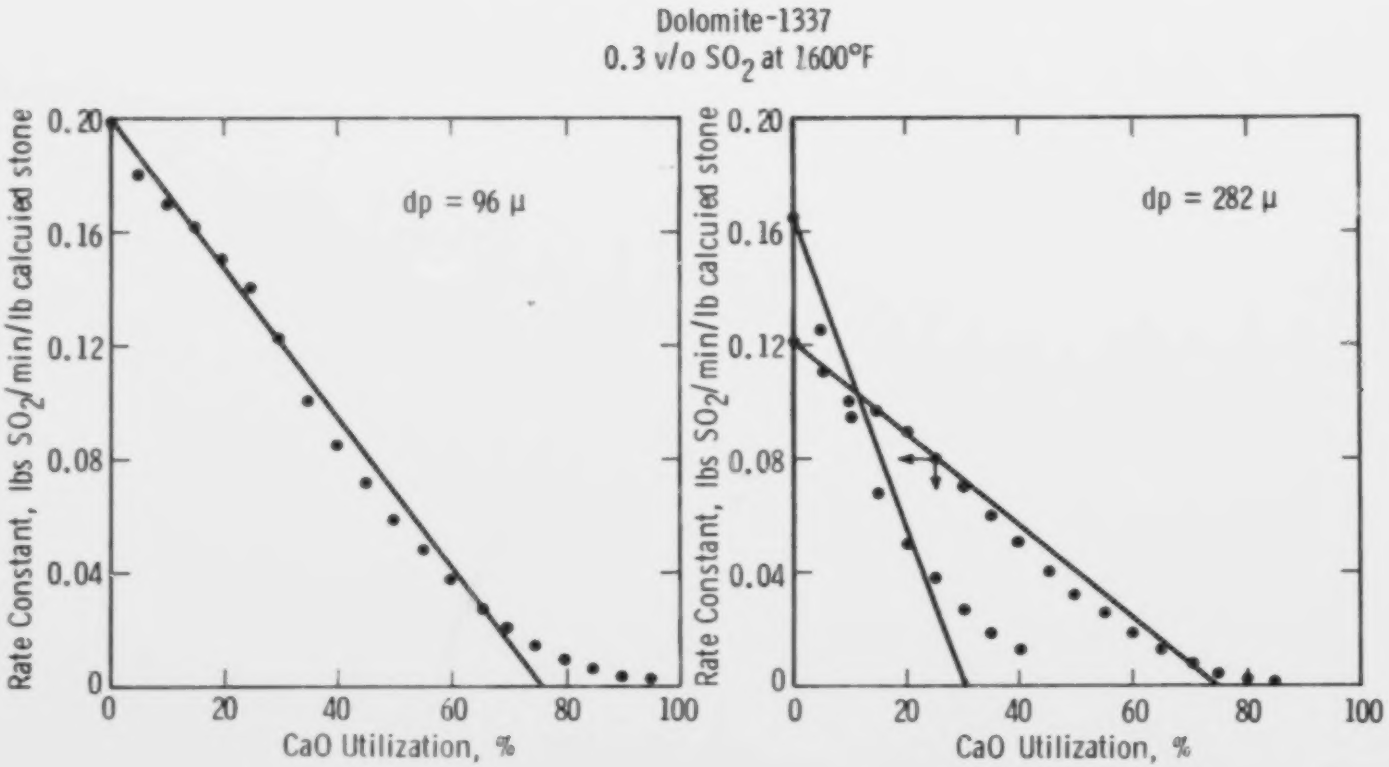

Fig. 1-12—Rate constants for Dolomite 1337 as a function of calcium oxide uitilization

Curve 645932-A

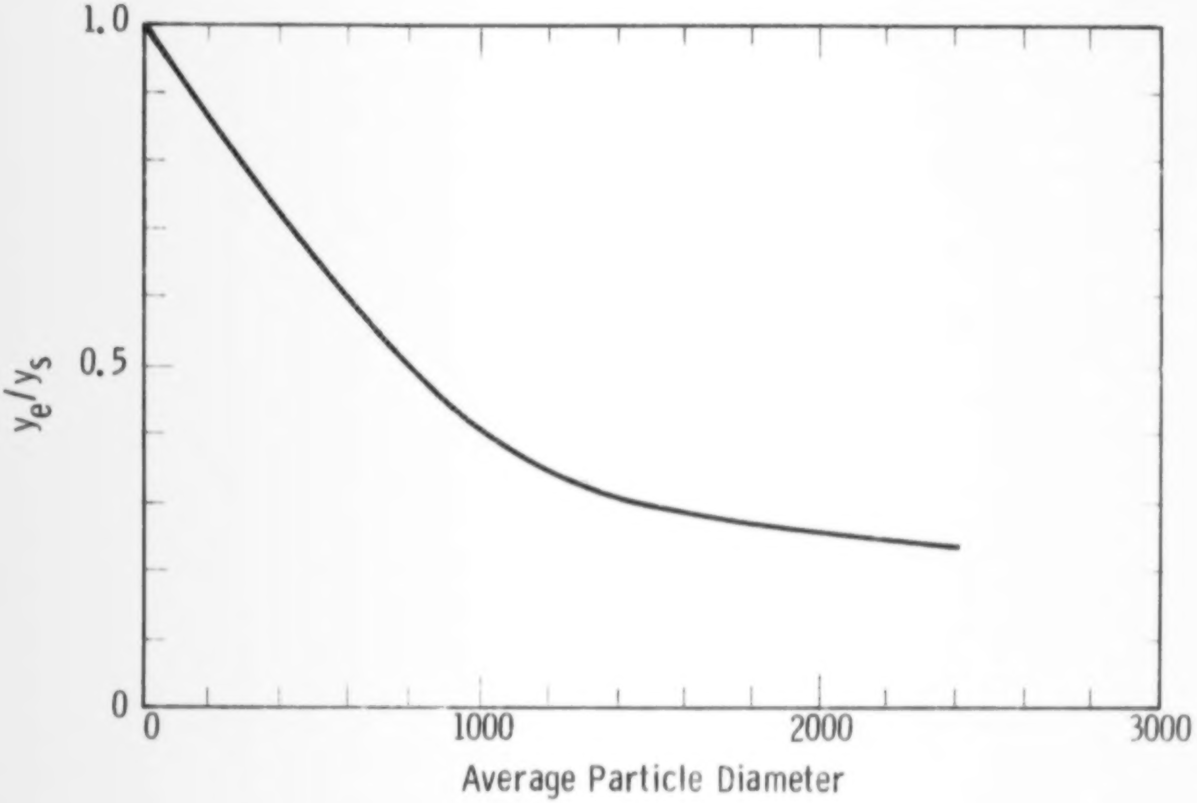

Fig. 1-13 – y_e/y_s vs particle diameter for limestone 1359

where

ϵ = contacting efficiency,

C_o = SO_2 concentration in the gas, and

C^* = effective equilibrium partial pressure of SO_2 at the stone surface due to locally reducing conditions.

If one assumes that ϵ and $\frac{C_o - C^*}{C_o}$ will be similar in the industrial boiler and in the Argonne experiment, then

$$H \text{ (industrial boiler)} = H_{exp} \times \frac{2.5 \text{ ft}}{1.25 \text{ ft}} \times \frac{3 \text{ ft/sec}}{12.5 \text{ ft/sec}} \times \frac{0.7}{0.9} = 4.25$$

$$r = 6 \times 0.7 = 4.2$$

$$R = 1/2(5.2 - \sqrt{(3.2)^2 + \frac{4 \times 4.2}{4.25}}) = 0.71.$$

An alternative computation for the same boiler could be based on PER data [80, FBC test #72]:

R = .347

dp = 840 μ (1359 limestone)

Ca/S = 2.6

1640°F

V = 13 ft/sec

18" static bed depth (assume 50% of bed as stone, 50% ash).

The experimental H is 1.8 and

$$H\text{(boiler)} = 1.8 \times \frac{2.5}{.5 \times 1.5} \times \frac{13}{12.5} \times \frac{7}{9} = 4.9$$

$$r = 4.2$$

$$R = 1/2(5.2 - \sqrt{(3.2)^2 + \frac{4 \times 4.2}{4.9}}) = 0.75 .$$

The closeness between the two results is encouraging, although it should be noted that the assumption of percentages of stone and ash in the bed were necessary in the second estimate since sufficient data were unavailable to calculate this quantity. The assumptions used for the model and the small-scale fluidized bed data give a conservative estimate of the SO_2 removal which can be expected for an industrial boiler.

Modification of Koppel's Model for Partially Sulfated Stone

Koppel's model is readily modified to account for partially sulfated stone being fed to the boiler for SO_2 absorption duty. This would occur physically when regenerated stone is recycled back to the boiler beds. Regenerated stone will be partially sulfated, especially if regeneration is accomplished in well-mixed, fluidized beds.

Koppel's equation for sulfur removal becomes

$$R = \frac{1}{1 - e^{-\alpha}} \left\{ 1 - \frac{\left[\alpha e^{-(1 - \frac{R}{r} - \frac{y_i}{y_e})H} - (1 - \frac{R}{r} - \frac{y_i}{y_e})He^{-\alpha}\right]}{\alpha - (1 - \frac{R}{r} - \frac{y_i}{y_e})H} \right\} \qquad (1\text{-}9)$$

where

y_i is the sulfur loading of boiler inlet sorbent (lbs of S/lb of CaO). For fresh stone $y_i = 0$. All other terms are defined on pages 9 and 10.

The average sulfur loading of sorbent leaving the boiler is given by

$$\overline{y} = y_i + \frac{FL_o R}{w} \qquad (1\text{-}10)$$

Figure 1-14 shows the variation of sulfur removal with Ca/S ratio (r) for different inlet sulfur loadings (y_i/y_e) and SO_2 generation profiles (α).

Curve 64 312-B

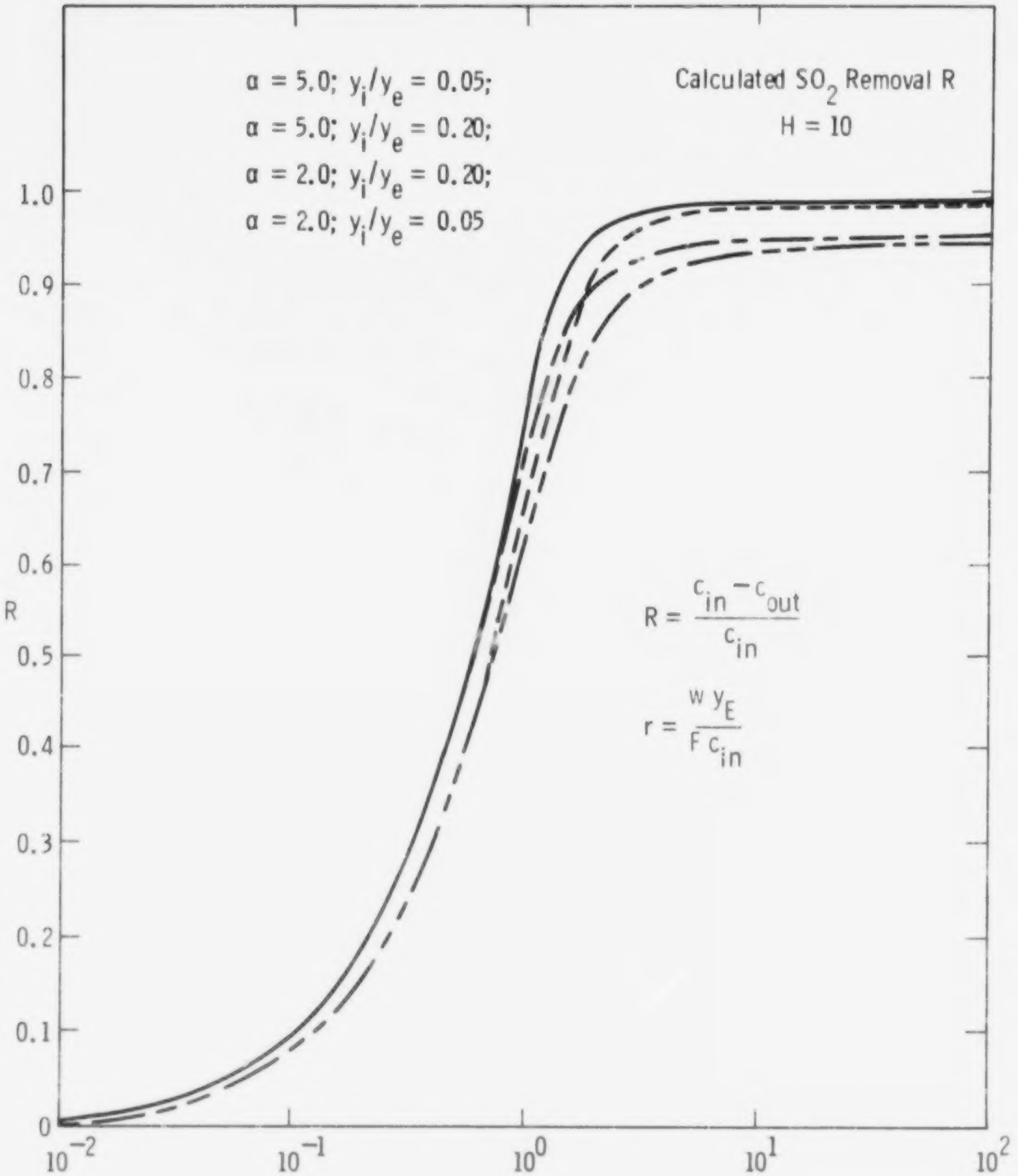

Fig. 1-14—Theoretical SO_2 Removal efficiency using partially sulfated stone

Deep Beds with Internals. In deep beds with internal heat transfer surface, circulation of solids may be impeded. If stone is fed to the top of such a bed and removed near the bottom, a staged countercurrent effect may be achieved. In such a case, the assumptions of well-mixed solids and uniform bed temperature made in deriving Koppel's model cannot be justified. In this case SO_2 removal may be computed by dividing the bed into a number of segments with each being considered as well mixed.

Sorbent Regeneration And Sulfur Recovery Systems

A number of processes have been proposed for converting a sulfated limestone or dolomite sorbent from a fluidized bed boiler back to an oxide or carbonate. Thus regenerated, the sorbent can be returned to the boiler for further capture of sulfur as SO_2 from the combustion gases. The sulfur emerges from the various proposed regeneration processes as either H_2S or SO_2 mixed with gaseous reactants and reaction products. Further processing of this mixture then recovers sulfur as a final product, either elemental S or sulfuric acid, H_2SO_4.

The optimum choice of a sorbent regeneration and sulfur recovery process is a technical and economic problem. Its solution depends on the operating conditions of the boiler, on the type of sorbent used, on the market for sulfur and/or sulfuric acid, and on a number of chemical kinetic, and operational factors which have not yet been thoroughly explored.

A listing and brief descriptions of some of the proposed sorbent regeneration processes follows.

1. Reduction of calcium sulfate to calcium sulfide and subsequent regeneration with steam and CO_2.

$$CaSO_4 + \begin{Bmatrix} 4\ H_2 \\ 4\ CO \end{Bmatrix} \rightarrow CaS + \begin{Bmatrix} 4\ H_2O \\ 4\ CO_2 \end{Bmatrix} \qquad (1\text{-}11)$$

$$CaS + H_2O + CO_2 \rightarrow CaCO_3 + H_2S \qquad (1\text{-}12)$$

This process has been investigated by Consolidation Coal and the City College of New York. To produce an effluent gas with a sufficiently high concentration of H_2S to allow it to be fed directly to a Claus plant for sulfur recovery requires: 1) sufficiently rapid kinetics for Equation 1-12 at 800°F so equilibrium between the gas and solids is approached; or 2) a CO_2-H_2O regenerating gas essentially free of nitrogen. Such a mixture could be produced from a steam-oxygen blown gasifier, a steam-methane reformer, or a hot carbonate flue gas scrubber. Alternatively, an air-blown gasifier could be used, and the regenerated H_2S gas concentrated with amine wash towers.

2. Direct reduction of calcium sulfate to calcium oxide and SO_2 employing partially oxidized fuel as the reducing agent.

$$CaSO_4 + \left\{ \begin{matrix} H_2 \\ CO \end{matrix} \right\} \rightarrow CaO + SO_2 + \left\{ \begin{matrix} H_2O \\ CO_2 \end{matrix} \right\} \qquad (1\text{-}13)$$

This process is thermodynamically favored at low pressure and high temperature. But Esso and Consolidation Coal have both shown that ash agglomeration problems may be encountered with this system if there is a significant ash fraction carried into the regenerator with the sorbent or with the fuel.

3. Reduction of calcium sulfate to calcium sulfide and subsequent oxidation to form calcium oxide and SO_2.

$$CaSO_4 + \left\{ \begin{matrix} 4\ H_2 \\ 4\ CO \end{matrix} \right\} \rightarrow CaS + \left\{ \begin{matrix} 4\ H_2O \\ 4\ CO_2 \end{matrix} \right\}$$

$$CaS + \frac{3}{2}\ O_2 \rightarrow CaO + SO_2 \qquad (1\text{-}14)$$

Consolidation Coal and Esso (England) have investigated this technique experimentally at atmospheric pressure with encouraging results. However, the equilibrium of the SO_2 rejection reaction is unfavorable for the pressurized boiler plant; the maximum SO_2 concentration is about 1% in the gases when regeneration temperatures lower than 2000°F are used.

4. Reduction of calcium sulfate to calcium sulfide and subsequent regeneration by reaction with CO_2. FMC Corporation has investigated the CO_2-calcium sulfide reaction at temperatures of 2000 to 2200°F. The temperatures used for this reaction appear too high to avoid ash agglomeration and absorbent deactivation problems.

5. Reduction of calcium sulfate to calcium sulfide and preparation of a calcium sulfide water slurry followed by low-temperature regeneration with CO_2.

Of these five processes, the first two appear most attractive in conjunction with fluid bed combustion boilers. They have been selected for preliminary analysis and design of the regenerator for both the high- and the low-pressure utility boiler operations.

Sulfate-Sulfide Reduction, Steam-CO_2 Reaction to Produce H_2S (Pressurized Operation)

This limestone/dolomite regeneration scheme incorporates a two-stage process in which the calcium sulfate is first reduced to calcium sulfide. The calcium sulfide is next converted to calcium carbonate evolving an H_2S-rich gas suitable for feed to a conventional Claus Plant. Squires [89] has outlined this process and reported some preliminary experimental results on it. The essential elements for this regeneration process are shown in Figure 1-15.

The gas producer is used to produce reducing gas for the first-stage regenerator. Coal, fuel oil, natural gas or char is oxidized by an air-steam mixture; steam controls reducing gas temperature and provides H_2 for reduction. Fuel to the gas producer could include char elutriated from the boiler beds. In this case the gas producer replaces the carbon burn-up cell.

Fuel combustion, gasification and the water gas shift reaction occur in the gas producer.

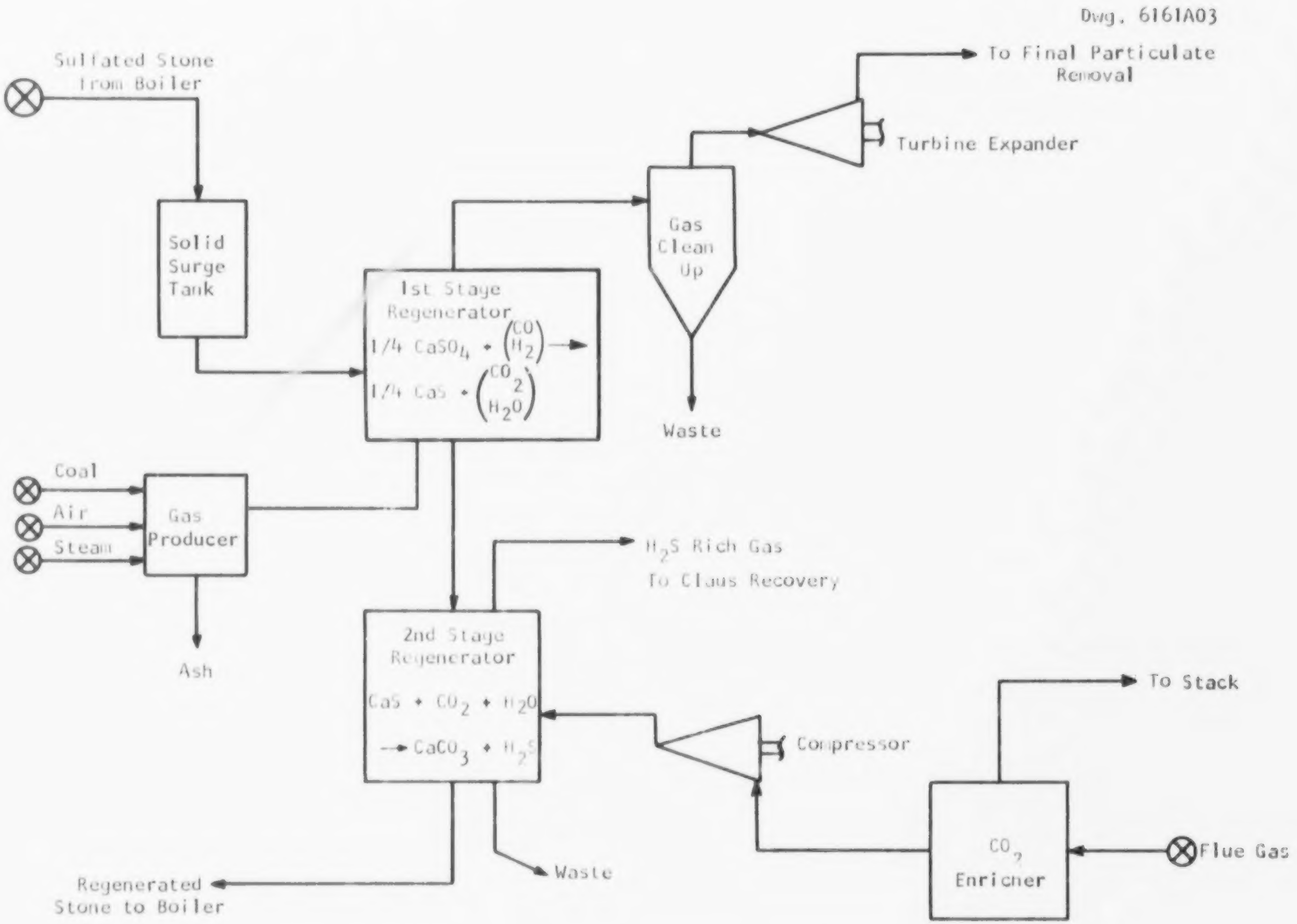

Fig. 1-15—Sulfate-sulfide reduction regeneration schematic

Fuel combustion

$$C + O_2 \rightarrow CO_2 \qquad (1\text{-}15)$$

$$H_2 + \frac{1}{2} O_2 \rightarrow H_2O \qquad (1\text{-}16)$$

Fuel gasification

$$C + \left\{\begin{matrix} H_2O \\ CO_2 \end{matrix}\right\} \rightarrow CO + \left\{\begin{matrix} H_2 \\ CO \end{matrix}\right\} \qquad (1\text{-}17)$$

Water gas shift

$$CO + H_2O \rightarrow CO_2 + H_2 \qquad (1\text{-}18)$$

In the first regenerator stage, calcium sulfate from the boiler is reduced to calcium sulfide. This regenerator reactor can be a bed fluidized by the reducing gases from the gas producer, a moving bed, or a panel bed. The choice has to be made on the basis of reaction rates and solids handling processes. Preliminary process designs have been based on a fluidized bed operating at about 1500°F and 8 atm.

The reduction reactions occurring in the first regenerator stage with a dolomite sorbent are

Sulfate reduction

$$[CaSO_4 + MgO] + \left\{\begin{matrix} 4\ H_2 \\ 4\ CO \end{matrix}\right\} \rightarrow [CaS + MgO] + \left\{\begin{matrix} 4\ H_2O \\ 4\ CO_2 \end{matrix}\right\} \qquad (1\text{-}19)$$

Carbonation

$$[CaO + MgO] + CO_2 \rightarrow [CaCO_3 + MgO] \qquad (1\text{-}20)$$

Water gas shift

$$CO + H_2O \rightarrow CO_2 + H_2$$

Some undesirable side reactions may also occur in the first regenerator stage.

Side reactions

$$CaSO_4 + \begin{Bmatrix} H_2 \\ CO \end{Bmatrix} \rightarrow CaO + \begin{Bmatrix} H_2O \\ CO_2 \end{Bmatrix} + SO_2 \qquad (1\text{-}21)$$

$$CaS + H_2O \rightarrow CaO + H_2S \qquad (1\text{-}22)$$

$$CaS + CO_2 \rightarrow CaO + COS \qquad (1\text{-}23)$$

$$CaS + \frac{1}{2} CO_2 \rightarrow CaO + \frac{1}{2} CS_2 \qquad (1\text{-}24)$$

$$\frac{1}{2} CaSO_4 + \frac{3}{2} CaS \rightarrow 2\ CaO + S_2 \qquad (1\text{-}25)$$

The sulfate reduction reactions are thermodynamically favored by low temperature and unaffected by pressure. Below 1600°F the equilibrium constants for these reactions are greater than 10^8 (CO_2/CO, $H_2O/H_2 > 100$); thermodynamically, these reactions virtually proceed to completion. In actuality, the degree of conversion of $CaSO_4$ will be limited either by the supply of reducing gas or by the kinetics of the reduction reaction.

Low temperature and high pressure suppress the equilibrium formation of SO_2, COS, H_2S, CS_2, and S_2 while enhancing recarbonation of calcium oxide. Recarbonation should tend to drive the reduction reactions by removing CO_2 from the gas.

One caution must be exercised in selecting the first-stage regenerator operating temperature and pressure. The steam partial pressure must be kept below the level at which $Ca(OH)_2$ will form; the binary system $Ca(OH)_2$-CaO forms a low melting eutectic at 1450°F, and $Ca(OH)_2$-$CaCO_3$ forms a low melting eutectic at 1180°F. Curran [21] reports that melts form at temperatures of 1500-1600°F whenever the steam pressure exceeds 13 atm. Within these temperature-pressure limitations, the actual operating temperature and pressure will depend on chemical kinetic, equilibrium, and overall process considerations.

Useful energy may be extracted from gas leaving the first-stage regenerator vessel. This gas is at high temperature, 1500°F, and at high pressure, if the regenerator is used in conjunction with a high-pressure boiler. Passing the gas through a turbine expander can recover some of this energy.

Reduced solids flow from the first-stage regenerator through a stand pipe leg into the second-stage regenerator vessel, where steam and carbon dioxide convert calcium sulfide to calcium carbonate. This vessel contains a fluidized bed operating at about 1100°F and 12 atm; it produces a gas with 23% H_2S on a dry basis. A hydroxide-carbonate, $Ca(OH)_2$-$CaCO_3$, eutectic melt may form at temperatures above 1180°F. Primary reactions occurring in the bed are:

H_2S formation

$$[CaS + MgO] + H_2O + CO_2 \rightarrow [CaCO_3 + MgO] + H_2S \qquad (1\text{-}26)$$

Carbonation

$$[CaO + MgO] + CO_2 \rightarrow [CaCO_3 + MgO] \qquad (1\text{-}27)$$

Increased H_2S concentration is thermodynamically favored by the combination of low temperature and high pressure, which also suppresses competing side reactions. Although an equal molar ratio of steam to CO_2 in the regenerator top gas yields the maximum theoretical H_2S concentration, regenerating with excess steam produces a higher H_2S concentration on a dry basis. CO_2 requirements are also reduced by this approach. A feed gas comprising primarily steam and CO_2 is required to obtain high H_2S concentration.

Feed gas for the second-stage regenerator is obtained by stripping CO_2 from boiler flue gas in a regenerable hot-carbonate or amine scrubber solution. A CO_2 enricher unit, commercially available, consists of two large towers -- one an absorber and the other a regenerator -- a reboiler, a solution cooler, and solution pumps. A cost estimate for a Benfield designed system is included in Appendix I . A compressor is required to provide the CO_2-H_2O reactants at the pressure of the second-stage regenerator.

Pure CO_2-steam mixtures can also be obtained from a steam-oxygen blown gasifier operating at elevated pressure. An on-site oxygen plant for such a regeneration process, however, would be an expensive alternative.

A CO_2-H_2O gas from an air-coal-steam combustor can also be used to generate H_2S. The H_2S concentration in the off gas would be much lower than when CO_2-H_2O undiluted by N_2 was employed. A selective solvent could then be employed, however, to enrich the H_2S concentration before further treatment.

The two-stage sulfate-sulfide sorbent regeneration process appears particularly adapted to high-pressure boilers. Although high pressure does not enhance the first-stage reactions, it decreases the formation of SO_2 and other undesirable byproducts. Vessel diameter for a particular gas flow rate is likewise reduced. High H_2S concentration in the acid gas leaving the second stage is also thermodynamically favored by high pressure. Compressing the steam and CO_2 feed to this reactor and the air to the first-stage reactor, however, increases capital and operating costs for the process.

Before entering the Claus plant for conversion to elemental sulfur, the H_2S-rich gas may be expanded to approximately 2 atm. In addition, Claus plant operation will be improved if steam is condensed from the acid gas prior to entering the Claus plant.

This pressurized process scheme, or reaction sequence, can also be used in conjunction with low-pressure boilers. Spent stone would be sent to a lock hopper either on leaving the boiler or between the first and second regeneration stages. Regenerated stone would likewise be held in a lock hopper before being recycled to the boiler. The remainder of the regeneration system will be the same as for the high-pressure boilers.

Should reaction kinetics for the second-stage reaction of calcium sulfide with steam and CO_2 be sufficiently rapid at 800°F, low-pressure operation of the second stage would be feasible. Negligible increased heat duty would be placed on the boiler even though stone would be removed at 1600°F and returned below 800°F.

Process Alternatives

A number of process alternatives can be considered for the sulfate-sulfide reduction regeneration process. These alternatives are concerned primarily with methods of modifying or eliminating the gas producer and with methods of handling gases supplied to or removed from the second-stage regeneration.

If natural gas is available, a steam-hydrocarbon reformer could be used in place of a gas producer to generate CO and H_2 for reduction. Only the reformer step would be required, eliminating the CO conversion and CO_2 removal steps.

Another possibility is to use natural gas directly in the first-stage regenerator for the reducing gas. There is some doubt that methane will react with hot sorbent particles to produce CO and H_2 or that natural gas will reduce $CaSO_4$. Esso has performed experiments using methane to reduce calcium sulfate to CaO while giving off SO_2. SO_2 concentrations produced, however, were very low, and it is not clear that methane is a suitable reductant to produce CaS at the lower temperature.

A solid or liquid fuel -- coal, char, or oil -- might also be added directly to the first-stage regenerator along with the $CaSO_4$ sorbent. The sorbent particles would be fluidized by steam and sufficient air to provide any heat required for regeneration. The CO_2 and H_2O gases from the first-stage regenerator would then be used in the second-stage regenerator. If larger quantities of CO_2 were required for H_2S production in this second stage, it would be provided by increasing the flow of $CaSO_4$ from the boiler to the first-stage regenerator. More carbonaceous fuel added to this stage would convert this $CaSO_4$ to CaS and the desired CO_2.

In the second-stage regenerator, only a fraction of the CaS would be converted to $CaCO_3$ to match the sulfur removal rate. The remainder of the CaS would be circulated back to the boiler where it would be reoxidized to $CaSO_4$. Solid sorbent, in this case, would serve a dual purpose, namely, as a sulfur removal agent and a coal oxidizer.

Experimental verification is required to prove this alternative process technically feasible. This process has possible problems connected with ash agglomeration, sorbent deactivation, and high solids flow rates. But it has the potential promise of eliminating a separate gas producer and CO_2 enricher systems.

Sulfur Recovery

Two main choices are available for the sulfur recovery step, depending on the final sulfur product desired -- elemental sulfur or sulfuric acid. A Claus plant would be the most reliable process for converting H_2S to sulfur. The split-flow process is usually used for gases containing less than 50% H_2S or with gas streams containing high concentrations of hydrocarbons. A three catalytic reactor, split-flow Claus plant will have a sulfur recovery efficiency of 92 to 94%, emitting a tail gas containing 7000 to 8000 ppm SO_2. Ford, Bacon and Davis has estimated the installed cost and utility requirements for such a plant. (See Appendix I.)

The Ralph M. Parsons Company suggests a different flow scheme for H_2S recovery. Their recommendation is presented below.(1)

> "If the H_2S-rich gas is produced, I would recommend a scheme which produces CO_2 for the H_2S generator by treating a flue gas with MEA or hot carbonate. I think it would be preferable to treat the gas from the H_2S generator at 10 atmospheres pressure with Purisol to extract both H_2S and CO_2 together, then regenerate the Purisol in two stages; the first stage would produce H_2S-rich feed gas for the Claus unit, and the second stage would produce CO_2 for recycling to the H_2S generator. This scheme would be more economical, I think, than the one Westinghouse proposes and I believe that it has a very substantial operating advantage in making the H_2S concentration in the Claus feed controllable at a level which would assure good operation of the Claus unit, regardless of variations in the sulfur content of the fuel."

Should this option be selected, the flue gas CO_2 scrubber system capacity would be reduced. In fact, this mode of operation could allow regeneration with a dilute CO_2-H_2O gas stream. The Claus plant cost should also be reduced.

An H_2S-rich gas stream is a suitable feed for a sulfuric acid plant. H_2S is first oxidized to SO_2, catalytically converted to SO_3, and absorbed in water. Browder [14] describes the Ralph M. Parsons Co., "Double Contact/Double Absorption" H_2S sulfuric acid plant for this service. A 15% H_2S gas stream when combusted with stoichiometric air will produce an 8.7% SO_2 gas stream. This is well within the feed limits for the Parsons DC/DA process.

Regardless of which sulfur recovery process is selected, the dust loading of gas entering the unit will be a key factor in the overall unit cost. The catalyst bed reactors cannot tolerate high dust loading, so a gas purification system may be required.

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(1) Personal communication, July 20, 1971.

Sulfate-Oxide Reduction to Produce SO_2 (Atmospheric Pressure Operation)

This limestone/dolomite regeneration scheme is a single-stage process in which the calcium sulfate from the fluidized bed boiler is converted to calcium oxide by treating it with reducing gases, CO and H_2. Sulfur is released from the process as SO_2 mixed with unreacted reductants and with CO_2 and H_2O oxidation products. The essential elements of the process are shown in Figure 1-16. Esso, Consol, Esso (England) and Pope Evans and Robbins have all reported experimental results with this process operating at atmospheric pressure and at a temperature in the range 1900 to 2000°F.

Esso [31] has sulfated batch beds of solid sorbent with simulated flue gas and with actual combustion gases burning high-sulfur coal in the bed. The bed of spent sorbent was then regenerated using reducing gas. The SO_2 concentration in the regenerator off gases was 8 mole % or greater at atmospheric pressure; gas-solid equilibrium was approached.

Esso (England) [29] performed similar experiments. They sulfated a batch bed of lime by burning high-sulfur fuel oil. Regeneration was accomplished by increasing the fuel rate, keeping the air rate constant. At 1050°C (1920°F) and one atm Esso was able to produce a 9% SO_2 gas from their regenerator.

Consol [106] continuously regenerated sulfated dolomite, using partial combustion of carbon monoxide gas at 1950°F. Regenerator off gas contained 3 to 7% SO_2. Dolomite was sulfated by combusting coal in a bed of dolomite. The fluidizing gas was enriched with SO_2 to provide the desired sulfur absorption rate. A continuous stream of dolomite was fed to and removed from both the regenerator and combustor.

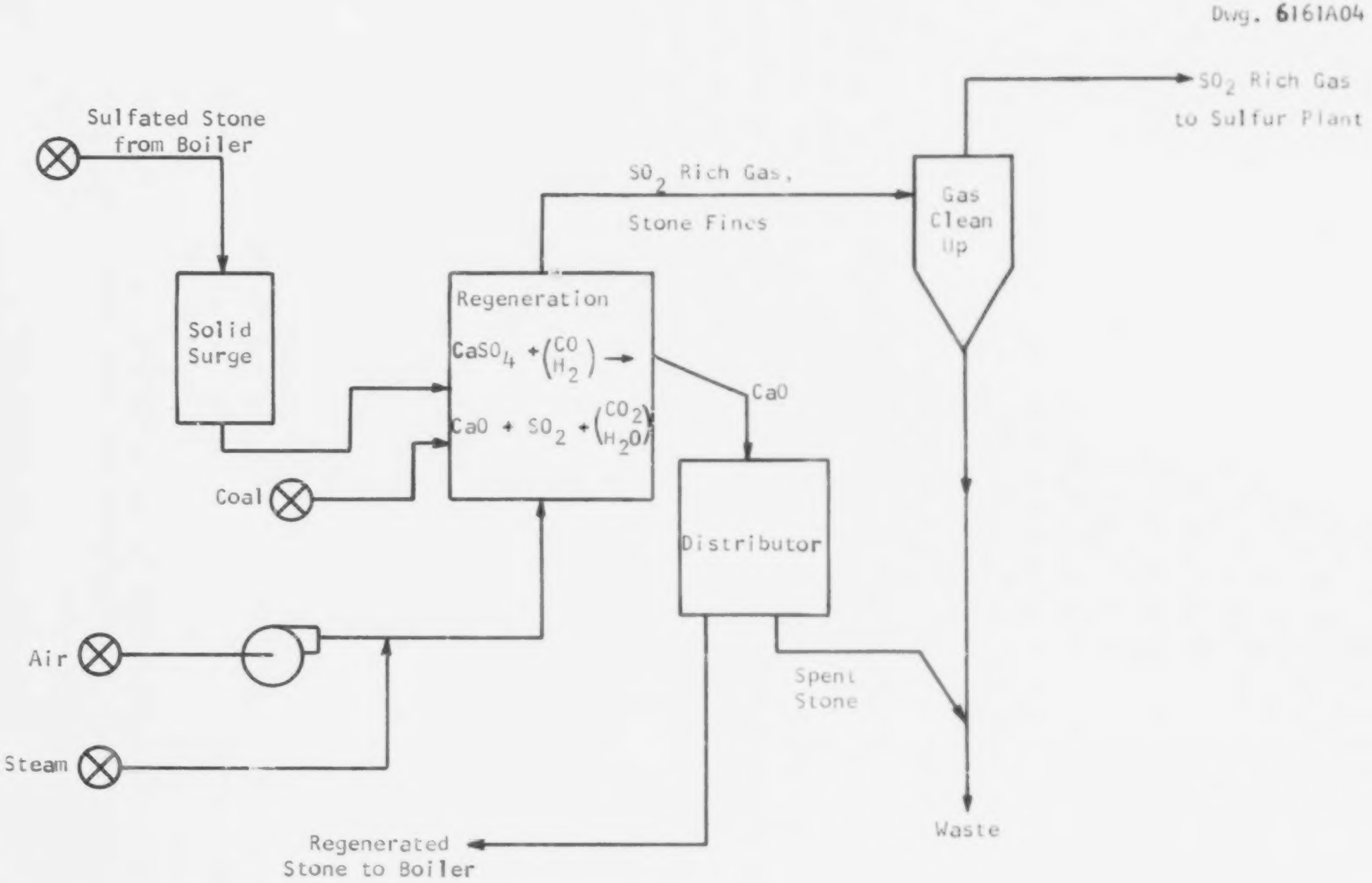

Dwg. 6161A04
Sulfated Stone from Boiler
Solid Surge
Coal
Air
Steam
Regeneration
$CaSO_4 + \binom{CO}{H_2} \rightarrow$
$CaO + SO_2 + \binom{CO_2}{H_2O}$
SO_2 Rich Gas, Stone Fines
Gas Clean Up
SO_2 Rich Gas to Sulfur Plant
CaO
Distributor
Spent Stone
Waste
Regenerated Stone to Boiler

Fig. 1-16—Sulfate-oxide reduction regeneration schematic

PER [75] conducted simultaneous combustion-regeneration experiments. They attempted to use their carbon burn-up cell for simultaneous regeneration and elutriated carbon burn-up. Although this dual functioning did not prove successful, they were able to maintain SO_2 levels of 3 to 4% in the CBC off gas by adding coal to the carbon burn-up cell. PER's current plans call for a three-reactor boiler system - a combustor, a regenerator, and a carbon burn-up cell.

Spent stone discharges from each bed of the fluidized bed boiler into a hold vessel -- one hold vessel per boiler module. The stone is then pneumatically transported to the regenerator vessel using air as the carrier gas.

The regenerator vessel is a fluid bed reactor operating at one to two atmospheres and 2000°F. Coal is charged into the reactor bed. An air-steam mixture fluidizes the reactor bed and combusts coal to provide reducing gas for reduction of the $CaSO_4$.

The primary regeneration reaction is

sorbent sulfate-oxide reduction

$$CaSO_4 + \left(\begin{matrix}CO\\ H_2\end{matrix}\right) \rightarrow CaO + \left(\begin{matrix}CO_2\\ H_2O\end{matrix}\right) + SO_2$$

The fuel combustion and gasification reactions also are carried out in the regenerator.

Fuel combustion

$$C + O_2 \rightarrow CO_2$$

$$H_2 + 1/2\ O_2 \rightarrow H_2O$$

Fuel gasification

$$C + \left\{\begin{matrix}H_2O\\ CO_2\end{matrix}\right\} \rightarrow CO + \left\{\begin{matrix}H_2\\ CO\end{matrix}\right\}$$

Only enough air is used to provide the necessary CO and H_2 reducing gases in the reactor. The reaction of air with the fuel also provides the heat required to carry out the high-temperature sulfate-oxide reduction process. Steam is added to the regenerator both to control the reactor temperature and to provide hydrogen for reduction. The exothermic heat of combustion must balance the endothermic requirements of reduction and gasification, sensible heat loss from solids circulation, and heat loss from the reactor.

Competing with the reduction to calcium oxide is the reduction to calcium sulfide according to the following reaction.

Sorbent sulfate-sulfide reduction

$$1/4\ CaSO_4 + \left(\begin{matrix}CO\\H_2\end{matrix}\right) \rightarrow 1/4\ CaS + \left(\begin{matrix}CO_2\\H_2O\end{matrix}\right)$$

This reaction does not produce SO_2 but it does consume large quantities of reducing gas for every mole of calcium sulfate reacted. This reaction can be minimized by judicious selection of reducing gas composition and reactor temperature and pressure. Some speculation is that sorbent is regenerated by first reducing some of the $CaSO_4$ to CaS according to reaction 1-19. The CaS then reacts with $CaSO_4$ according to reaction below to form CaO and drive off SO_2. This would be a solid state reaction.

Sorbent sulfate-sulfide reaction

$$3/4\ CaSO_4 + 1/4\ CaS \rightarrow CaO + SO_2 \qquad (1\text{-}28)$$

Chemical equilibrium in this reaction can establish the maximum concentration of SO_2 which can be attained in gases from the regenerator. A high percentage of SO_2 is favored by high operating temperatures -- around 2000°F, and low operating pressure -- about 1 atm. The SO_2 fraction in the stream to the sulfur recovery plant can be increased from 8-10% to 90% by volume by a dimethylamine, DMA, absorber-stripper process operating on the regenerator top gases subsequent to their cleanup and cooling.

Regenerated sorbent flows from the reactor vessel to a distributor vessel. Twenty-four lines return regenerated stone from the distributor to each bed in the boiler modules.

The problems involved in designing this sulfate-oxide sorbent regeneration process are selecting gas composition and flows and designating temperatures and pressures to

- Obtain high SO_2 concentrations (6 to 10% volume)
- Prevent production of CaS
- Provide the heat required by the endothermic, SO_2 -producing reactions and by the process of heating the sorbent.

The size of the regenerator vessel must also be selected so that combustion-gasification of the fuel and conversion of the sorbent sulfate to oxide can be carried out to the extent required by the sulfur sorption rate in the boiler.

This single-step regeneration process is most effective with low-pressure boiler operation. At high pressures the SO_2 concentrations produced appear too low for effective recovery of the sulfur. A high-pressure boiler operation would entail the use of high-temperature lock hoppers to step the stone down from boiler pressure to regenerator pressure and then to step it back up to boiler pressure.

Process Alternatives

Variations in the sulfate-oxide reduction regeneration process are concerned with alternative methods of providing reducing gases. Adding coal or char directly to the regenerator brings quantities of ash in contact with the sorbent. Such contact may cause problems of agglomeration and sorbent deactivation. A separate gasifier followed by an ash particle removal system can be used to furnish a reducing gas to the regenerator.

Sulfur Recovery

The SO_2-rich gas stream from the regenerator can be converted to sulfuric acid or elemental sulfur. A suitable acid production plant is commercially available at this time. Appendix I contains capital cost and operating cost estimates for sulfuric acid plants to treat the SO_2-rich regenerator gas. The sources of these estimates are Monsanto Enviro-Chem Systems Inc. and the Ralph M. Parsons Company.

A key design factor in any sulfur recovery process is the dust loading of the gas fed to the sulfur recovery plant. High dust loading can cause rapid catalyst failure. Approximately half the capital cost for Parsons' sulfuric acid plant is in the gas cleaning section of the plant. Monsanto specifies a maximum dust loading of 0.5 gr/scf for the gas entering the purification section.

No commercial process is readily available to convert the SO_2 to elemental sulfur. Allied Chemical [2] has investigated several direct reduction processes, including one developed by Asarco, to reduce gases containing 0.3%, 3.0-16%, and 90% SO_2. No process appeared practical for the 0.3% case. Although gases containing 3 to 16% SO_2 can be reduced directly, using coke or methane as a fuel, Allied concluded that enriching the SO_2 concentration up to 90% and then reducing the SO_2 is more economical and reliable. This conclusion is shared by the Ralph M. Parsons Co. and Ford, Bacon and Davis. (See Appendix I). A selective solvent, such as DMA, to absorb SO_2 and produce an enriched SO_2 gas stream is currently being developed. Again gas cleanliness -- less than

0.2 gr/scf particulates -- is important to the technical feasibility of this process.

This absorption-enrichment reduction scheme appears promising, particularly in conjunction with a high-pressure regenerator, provided that the SO_2 concentration in the regenerator remains above 3%. The sulfur recovery plant will be less susceptible to boiler and stone regenerator fluctuations since it will be fed a concentrated SO_2 stream. For the technical and economic feasibility estimations, sulfur recovery was carried out by the Asarco process, as described by Allied Chemical, using natural gas as a fuel.

Thermodynamic Data for Various Regeneration Processes

Equilibrium constants for the various regenerator reactions are given in this section. Data points referenced as "Squires" were obtained from [89]. Data points referenced as "Esso" were calculated from free energy data in [32]. Similarly, "Consol" refers to [21] and "Esso (England)" to [29].

Parenthesis, "( )", is defined as mole fraction, i.e., "(H_2S)" = mole fraction H_2S. "π" is total pressure in atmospheres. Equilibrium constants are either dimensionless or given in units of atmospheres.

Also presented graphically is the enthalpy above a 77°F datum for the solid and gaseous species. Solids data were obtained from Kubaschewski and Evans [59], and Strassburger [94]. Gaseous data were obtained from JANAF [46].

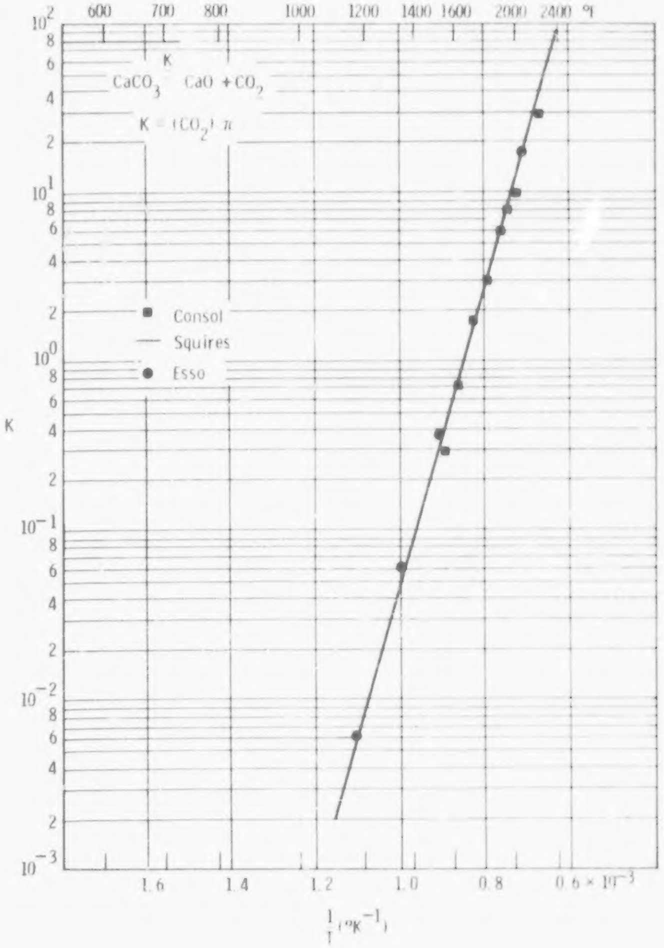


Fig. 1-17

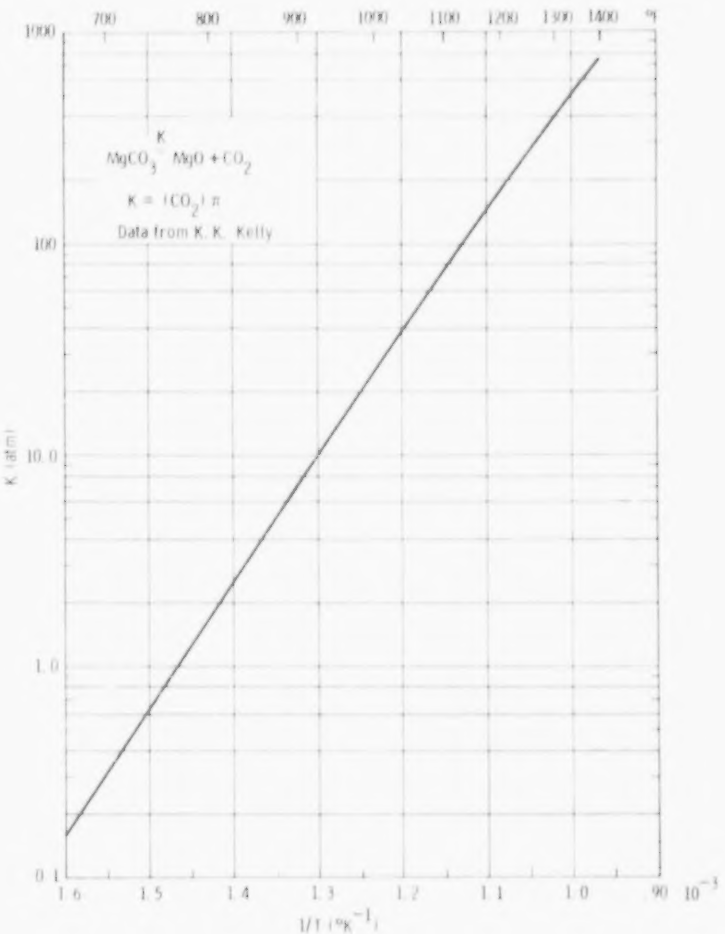
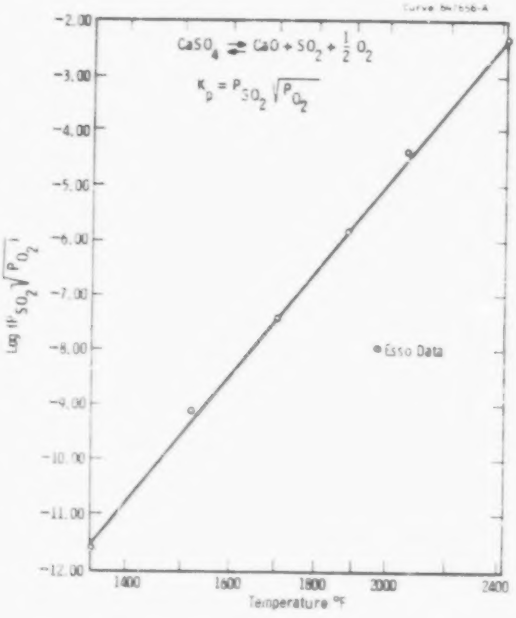


Fig. 1-18


Curve 647656-A
$CaSO_4 \rightleftharpoons CaO + SO_2 + \frac{1}{2} O_2$
$K_p = P_{SO_2} \sqrt{P_{O_2}}$
$\log (P_{SO_2} \sqrt{P_{O_2}})$
-2.00
-3.00
-4.00
-5.00
-6.00
-7.00
-8.00
-9.00
-10.00
-11.00
-12.00
● Esso Data
1400
1600
1800
2000
2400
Temperature °F

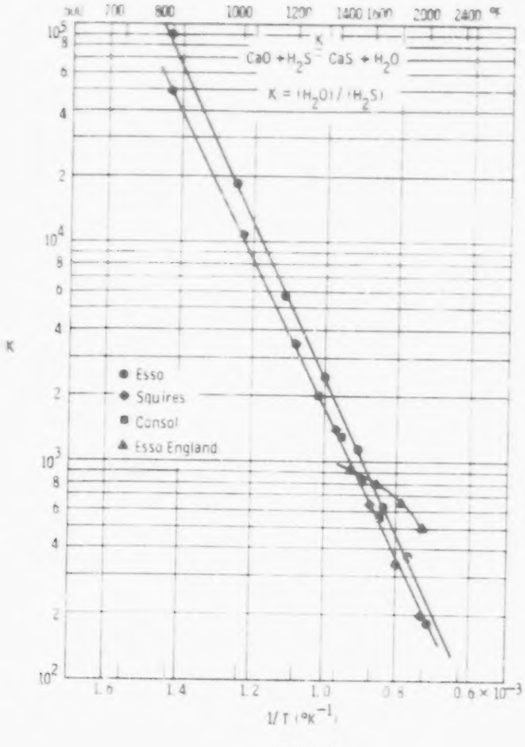

600
700
800
1000
1200
1400
1600
2000
2400 °F
$CaO + H_2S \overset{K}{=} CaS + H_2O$
$K = (H_2O) / (H_2S)$
K
10^5
10^4
10^3
10^2
8
6
4
2
● Esso
◆ Squires
■ Consol
▲ Esso England
1.6
1.4
1.2
1.0
0.8
0.6×10^{-3}
$1/T$ (°K^{-1})

Fig. 1–19

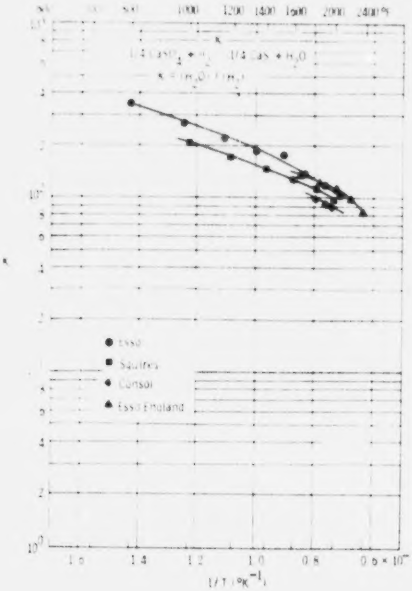


Fig. 1-20

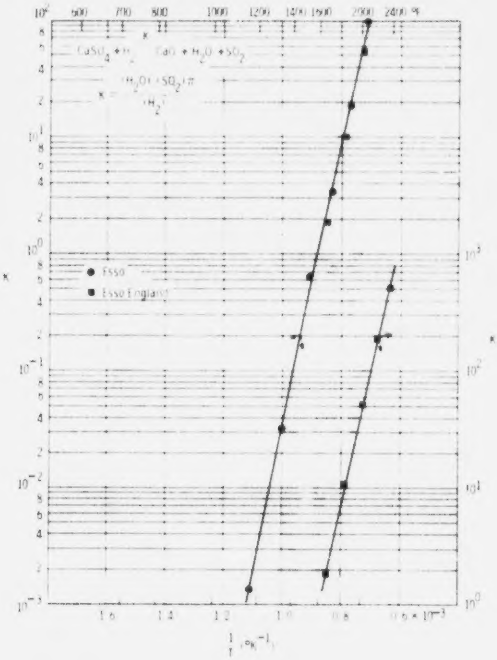


Fig. 1-21

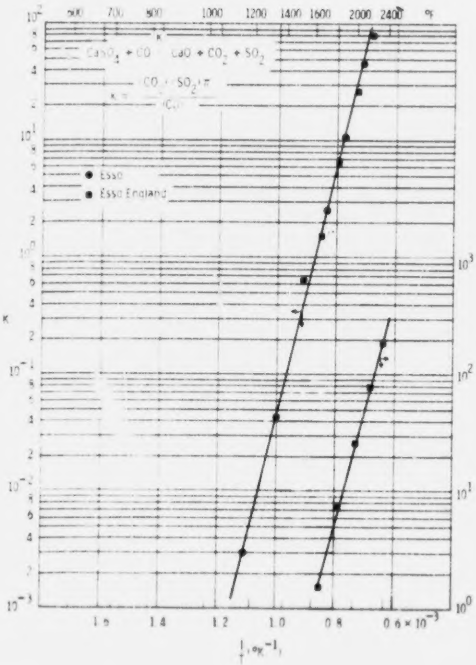


Fig. 1-22

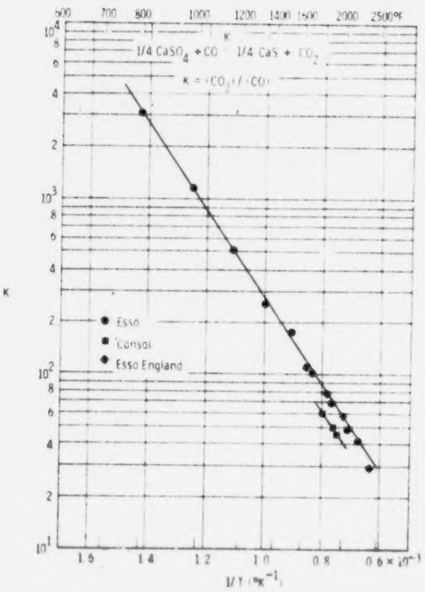


Fig. 1-23

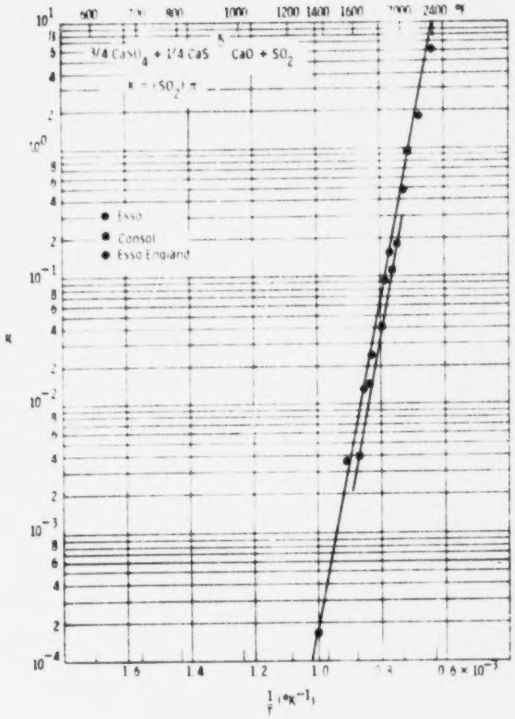


Fig. 1-24

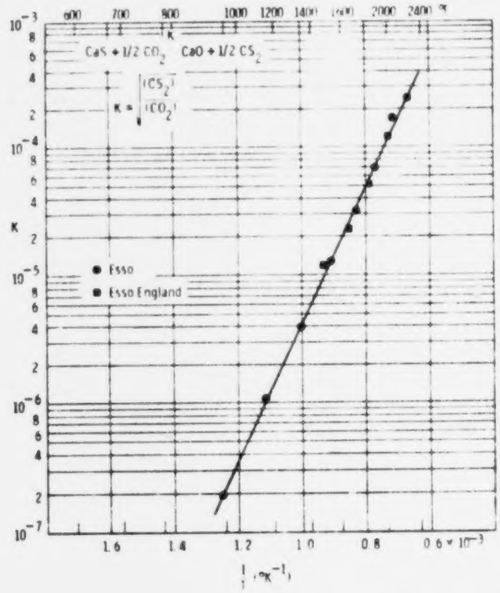


Fig. 1-25

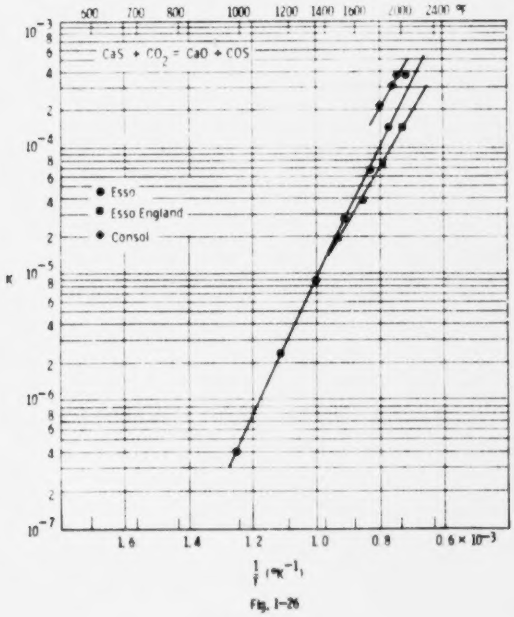


Fig. 1-26

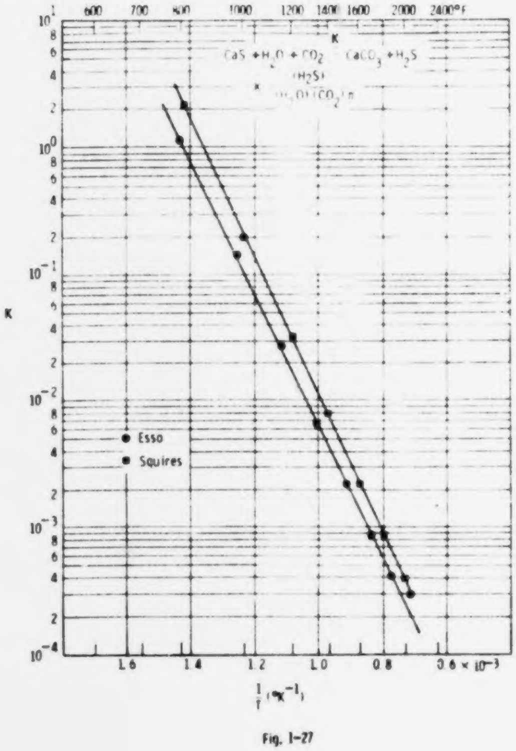


Fig. 1-27

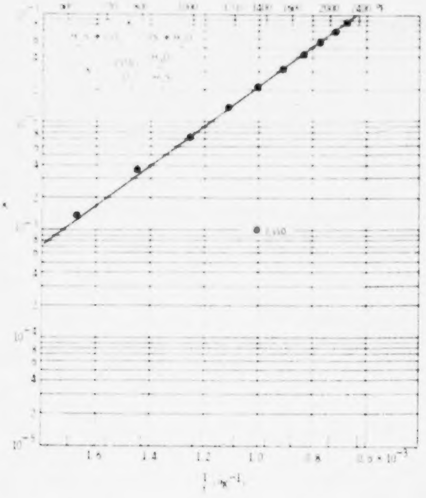


Fig. 1-28

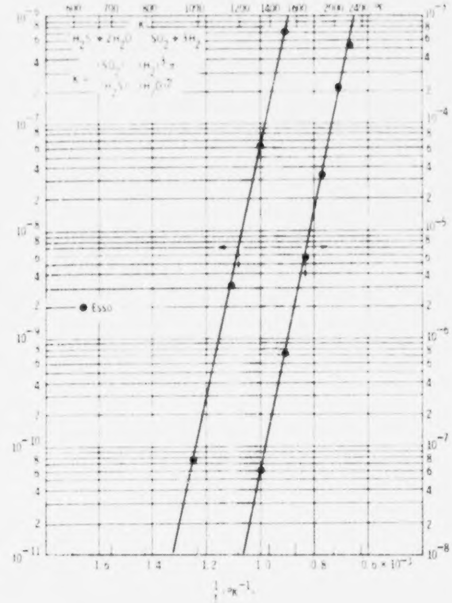


Fig. 1-29

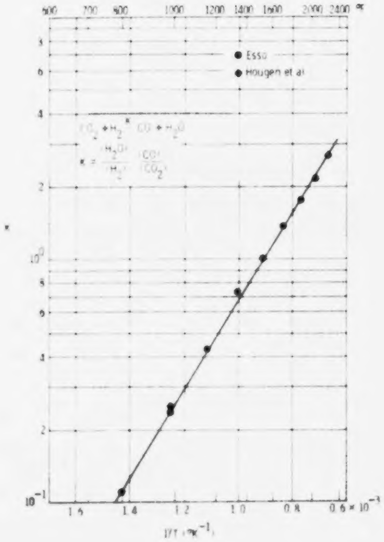


Fig. 1-30

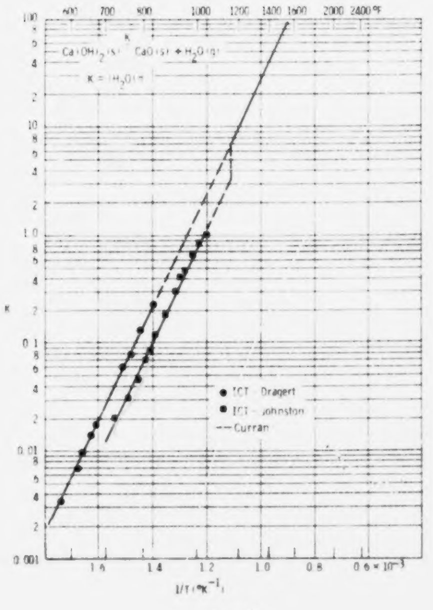


Fig. 1-31

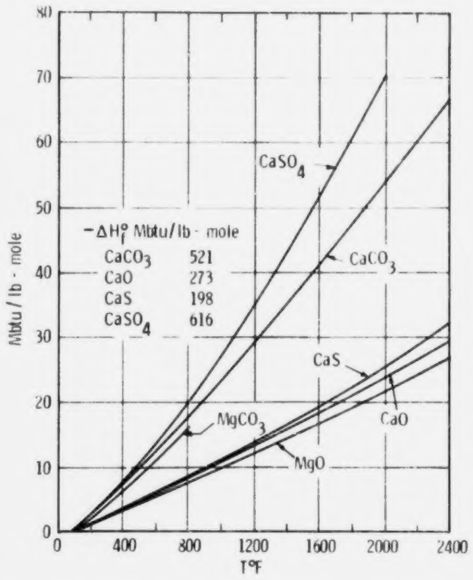
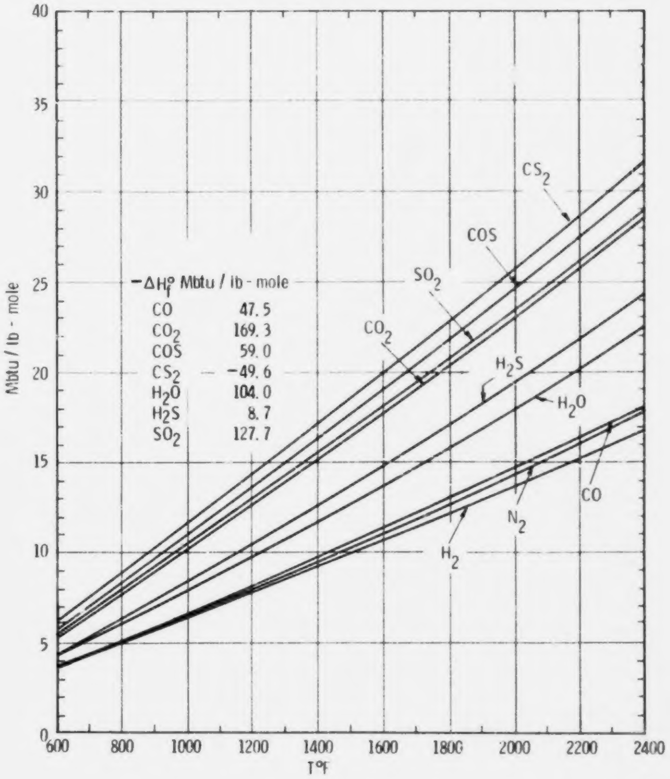


Fig. 1–32—Heat content (enthalpy) above 77°F datum mbtu / lb - mole
(Kubaschewski & Evans, Strassburger)



Heat content (enthalpy) above 77°F datum mbtu / lb - mole (JANAF thermochemical data)

Fig. 1–33

Minimization of NO_x Emissions

Nitrogen oxides (NO_x), predominantly nitric oxide (NO) with lesser quantities of NO_2, react photochemically in the atmosphere to generate smog and noxious compounds which have undesirable effects on plant and animal life. These gases are, therefore, air pollutants whose emission must be minimized.

NO_x can be formed in the combustion of coal and oil either by fixation of atmospheric nitrogen or by the oxidation of the nitrogen content of the fuel. At temperatures above 2700°F the N_2 of the air reacts directly with excess O_2 remaining after combustion to form NO

$$N_2 + O_2 \rightleftarrows 2NO^{(1)} \qquad (1\text{-}29)$$

In a conventional high-temperature combustion process, the quantity of NO formed can usually be estimated from equations involving the kinetics and the equilibrium of this fixation reaction. At temperatures below 2400°F, however, the rate of the reaction forming NO from N_2 and O_2 is very slow; the equilibrium quantity of NO which can be formed is also small. In a low-temperature fluidized bed combustion process, therefore, NO is formed as a combustion product from nitrogen contained in the fuel.

No way has been developed to predict the percent of fuel nitrogen converted to NO_x, but preliminary experiments indicate that it may be 20 to 40%. Concentrations of NO higher than those predicted by equilibrium of the fixation reaction can occur. Once formed, the NO can decompose by either of two reactions

$$2NO + 2CO \rightarrow N_2 + 2CO_2 \qquad (1\text{-}30)$$

$$2NO + 2SO_2 \rightarrow N_2 + 2SO_3 \qquad (1\text{-}31)$$

which may occur in the gas phase or, more likely, on solid surfaces in the combustor.

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(1) Free energy changes for the reaction can be obtained from the JANAF Tables [46].

NO_x Emissions From Conventional Boilers

The concentration of NO_x in stack gases from conventional power plants depends on the power output of the plant, the design of the boiler, and the nitrogen content of the fuel. Emission levels of 400 to 700 ppm in gas- and oil-fired plants [87, 92] and of 500 to 1000 ppm in coal-fired plants [68] have been measured in installations of 200 to 500 megawatts. Techniques such as low excess air, two-stage combustion, and flue gas recirculation can be used individually and in combination to reduce the NO_x emissions by 50% or more.(1) Emissions of 42 ppm from a 315 MW plant are predicted for boiler operation with flue gas recirculation and sub-stoichiometric operation at the burners with overfire air. [88]

NO_x Emissions From Fluidized Bed Combustors

Measured NO_x concentrations from pilot fluidized bed combustors operating at 1450-1750°F and 1 atm range from 250-1200 ppm depending on the operating conditions and coal size [4, 32, 69, 70, 71, 75, 76]. The highest NO_x concentration occurred with pulverized coal. At 1500°F and elevated pressure (4-6 atm), NO_x emissions decreased to 50-150 ppm, with higher values corresponding to larger quantities of excess air.

It might be postulated that the temperature in the neighborhood of a burning coal particle within a fluidized bed is sufficiently high, above 2400°F, to promote the fixation reaction (1-29) above. There is no indication, however, that such high temperatures are reached. Experiments [4] indicate rather that NO_x is produced by oxidation of the nitrogen in the fuel rather than by fixation. The effect of pressure and air quantity on NO_x emissions can then be explained by the effect of these operating variables on the rate of equation (1-29) above.

A variety of experiments have been carried out to determine the effects of various operating conditions on NO_x emissions from fluidized beds and to clarify the mechanisms involved in NO_x formation and reduction. The results of these experiments and possible interpretations are described in the following sections.

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(1) Bartok, W., A. R. Crawford, A. R. Cunningham, H. J. Hall, E. H. Manny, and A. Skopp, "Systems Study of NO_x Control Methods For Stationary Sources," Final Report to National Air Pollution Control Administration, GR-2-NOS-69, Contract No. PH-22-68-55, November 20, 1969.

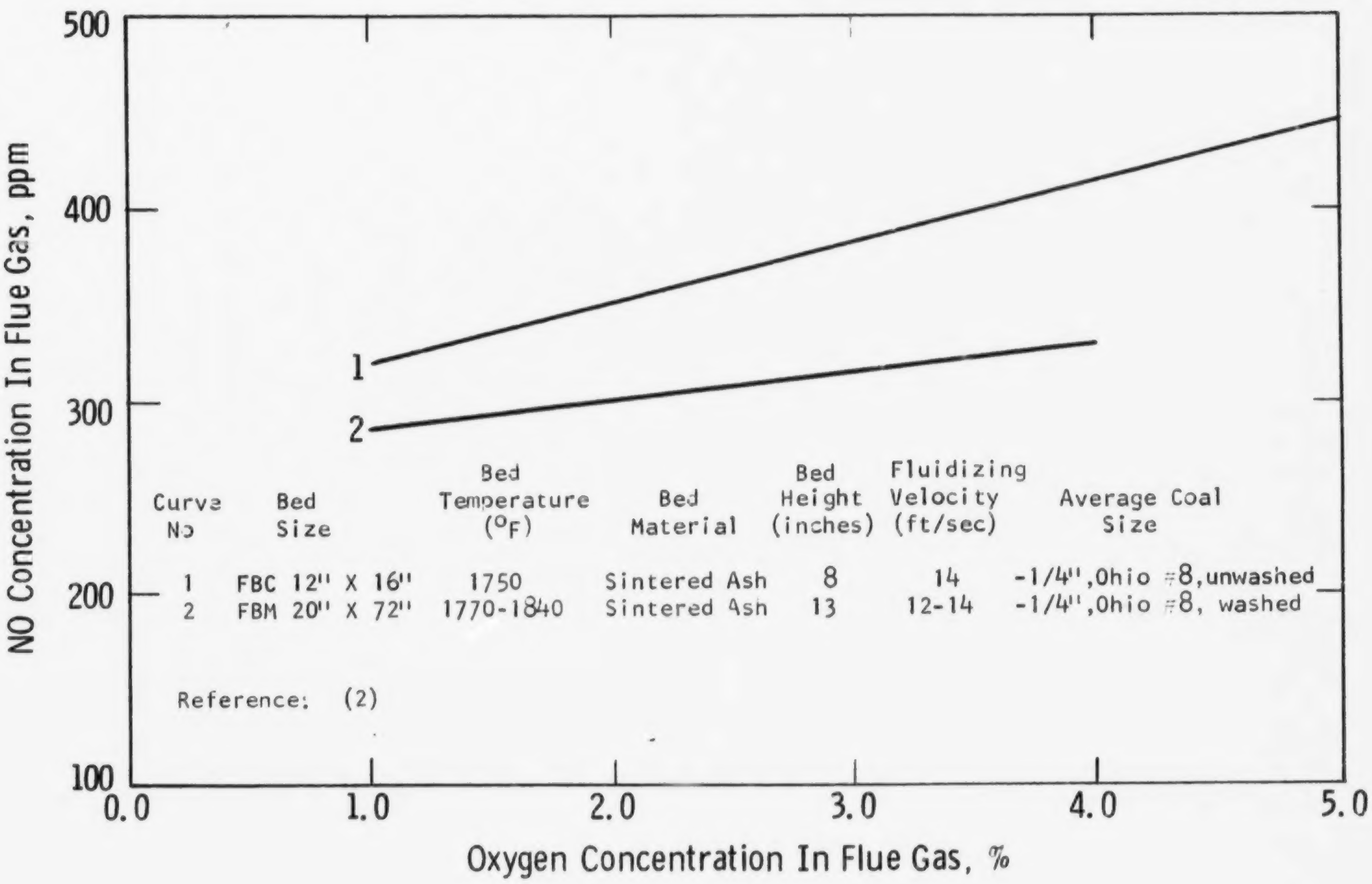
Combustion Gas Composition. Combustion gas composition has been modified in two primary ways: argon has been substituted for nitrogen in the combustion air to discover the source of the N in the NO_x emissions; the air/fuel ratio has been varied to determine the effect of excess oxygen on NO_x emissions. Argonne conducted a simple and direct demonstration that the nitrogenous content of the coal is the source of NO.[4] Argon, an inert gas, was substituted for the active component of nitrogen in the fluidizing gas. Since this substitution did not significantly affect the NO concentration in the flue gas, most of the NO_x obviously originated from the nitrogenous content of the coal. Additional evidence was obtained by combustion of natural gas, CH_4, also conducted by Argonne National Laboratory[4]. In these experiments, in which the only source of nitrogen was the fluidizing air, the observed NO_x concentration in the flue gas was less than 100 ppm, a value predicted by the nitrogen fixation equilibrium.

From the accumulated experimental evidence, there is no doubt that the nitrogenous content of the fuel plays a key role in NO_x generation in fluidized bed combustion. The amount of nitrogenous components converted to NO may depend on the temperature of combusting particles and the rate of combustion, which in turn depends on operating conditions such as bed temperature, excess air, fluidizing velocity, SO_2 sorbent injection, and coal particle size. These same factors also may influence the subsequent reduction of the NO_x formed to N_2 and combustion products.

Emission of NO from the fluidized bed combustor was found to increase with O_2 content in the flue gas as determined by the excess air rate. The correlations obtained by PER in their FBC and FBM test series[75] are presented in Figure 1-34.

The lower NO emissions at low excess air are believed to result from the reduction of NO by CO, which is more abundant during operation with low excess air. Substoichiometric operation gave even

Curve 643508-A



Curve No	Bed Size	Bed Temperature (°F)	Bed Material	Bed Height (inches)	Fluidizing Velocity (ft/sec)	Average Coal Size
1	FBC 12" X 16"	1750	Sintered Ash	8	14	-1/4",Ohio #8,unwashed
2	FBM 20" X 72"	1770-1840	Sintered Ash	13	12-14	-1/4",Ohio #8, washed

Reference: (2)

Fig. 1-34 –Effect of oxygen concentration on NO generation

lower NO_x emissions: at 68% stoichiometric air Esso found zero NO_x emissions. Esso was unable to find any correlation with excess air, however, because of scattered data [37].

Bed Temperature. Both Esso and Argonne observed a bed temperature effect on NO_x emissions. The experimental data are summarized in Figure 1-35. Decreasing bed temperature decreased NO_x emissions. It is not clear whether this temperature affects primarily the conversion of fuel nitrogen to NO_x or the reduction of the NO_x after its formation.

PER found no correlation between NO concentration and bed temperature. This contradictory observation is not surprising, because the measured NO concentrations are well above those predicted by thermodynamic equilibrium.

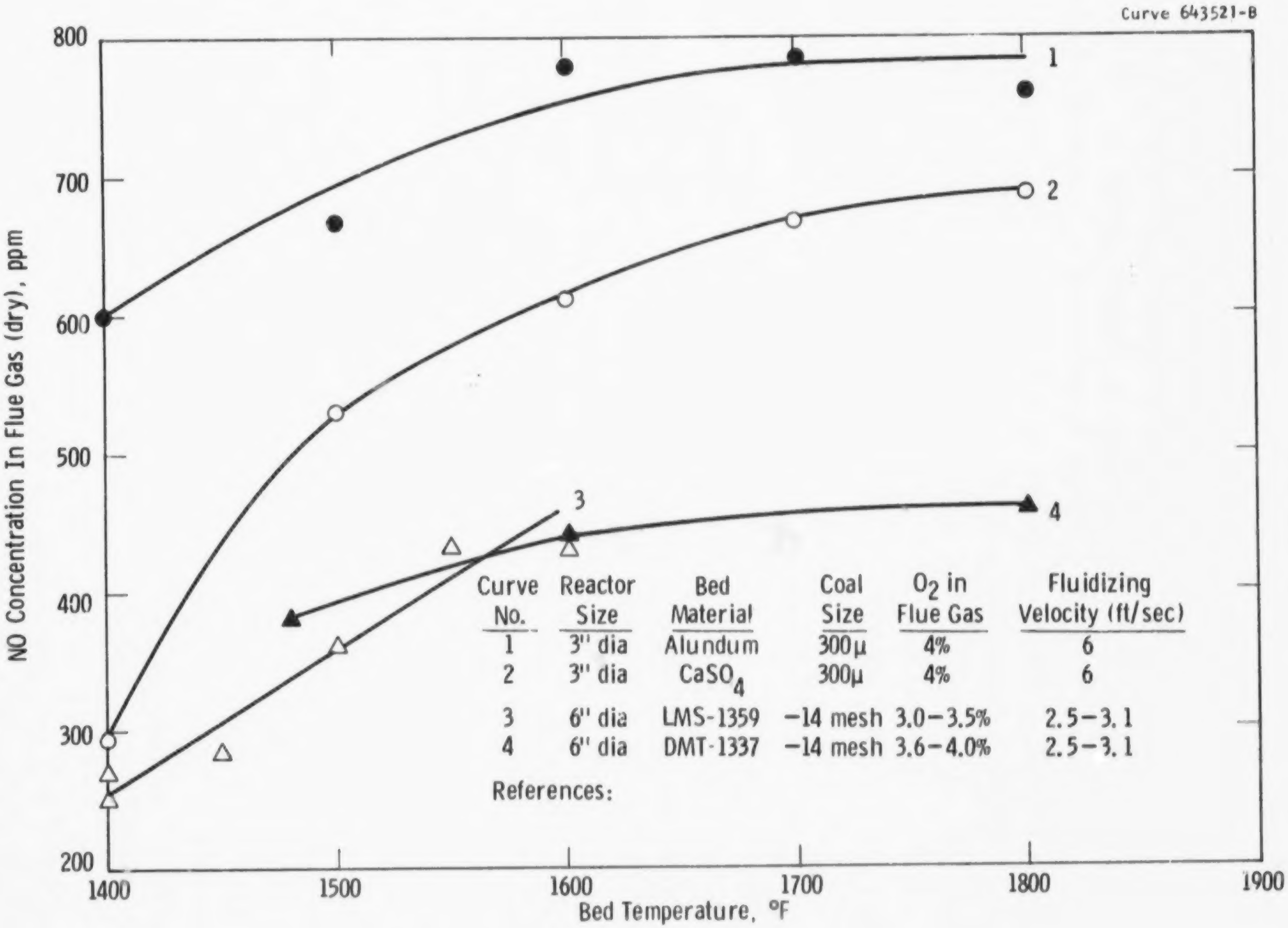
Coal Size. Coal size plots of available data (Figures 1-34 and 1-35) indicate that the NO_x emissions from fluidized bed combustion are dependent upon the coal particle size. At a bed temperature of 1600°F and with limestone or dolomite as bed material, PER [75, 76] reported a NO concentration in flue gas of 300-360 ppm using - 1/4" coal particles at 8-10 ft/sec fluidizing velocity; Argonne [4] obtained 380-500 ppm using - 14 mesh coal particles at 3 ft/sec; and Esso [32] measured 700-740 ppm using pulverized coal particles of size $\sim 200\mu$ at 3 ft/sec. These results may be somewhat confounded because Esso used coal with a higher nitrogen content (1.4 wt %) than Argonne's (1.1 wt %). But observations seem to confirm that smaller coal feed particles cause an increase in NO_x emissions.

Superficial Gas Velocity. At a given bed temperature, Esso found no effect of superficial fluidizing velocity on NO_x, while Argonne showed that the degree of reduction in NO concentration was inversely proportional to fluidizing velocity [4]. The conflicting results may be due to the fact that different coal sizes and reactor sizes were used. If the degree of NO decomposition controls the final level of NO_x concentration in the flue gas, it is reasonable to expect the NO_x reduction to be directly proportional to the gas residence time and thus inversely proportional to the superficial fluidizing velocity.

Bed Material. Esso observed consistently lower NO_x emissions from $CaSO_4$ than from alundum beds (see Figure 1-35). This fact was presented as evidence that a heterogeneous reaction between NO and CO might occur over $CaSO_4$, but not over the inert alundum. Argonne also measured low NO_x emissions from beds of limestone. A reduction of NO_x concentration in the flue gas up to 40% of its original value simultaneously with the reduction of SO_2 concentration upon addition of SO_2 sorbents to the fluidized bed was observed [4]. However, PER concluded that in general, NO_x emissions were not affected by addition of sulfur control sorbents.

The use of sorbents as bed material also decreases the amount of SO_2 in the fluidizing gases. In turn, NO_x emissions may be increased due to the effect of decreased SO_2 concentrations on the NO reduction reaction (1-31) above. Both Esso, operating at 1 atm, and the NCB [69, 70, 71] operating at 3.5 atm, reported that NO_x emissions decreased with increases in SO_2 concentration in flue gas. In batch operation carried out by Esso, much higher NO_x (1000-1200 ppm) was observed with the fresh stone when SO_2 removal was essentially complete. In the continuous long duration operation at NCB, the NO concentration was lowest (37 ppm) at the beginning of the test when the SO_2 concentration was at a maximum. For most of the test the concentration was 100-120 ppm, a value substantially lower than that obtained in atmospheric operation.

But it also appears that the sorbent catalyzes the reduction reaction (1-31). At 1630°F, Esso found that NO and SO_2 did not react to an



Curve No.	Reactor Size	Bed Material	Coal Size	O_2 in Flue Gas	Fluidizing Velocity (ft/sec)
1	3" dia	Alundum	300μ	4%	6
2	3" dia	$CaSO_4$	300μ	4%	6
3	6" dia	LMS-1359	−14 mesh	3.0−3.5%	2.5−3.1
4	6" dia	DMT-1337	−14 mesh	3.6−4.0%	2.5−3.1

Fig. 1-35 -Temperature effect on NO generation

appreciable extent in gas phase with gas residence time of about one second. However, when SO_2 was injected into a fluid bed of partially sulfated stone with N_2 as a transport gas, NO_x emissions were reduced. Possible mechanisms for this process have been postulated by Esso [32], but these mechanisms have not been confirmed.

Conclusions

The experimental observations on NO_x emissions from fluidized beds indicate that NO is formed from fuel nitrogen in the burning process. This NO can subsequently be reduced in the bed by CO and/or SO_2 according to reactions (1-30 and 1-31), which may be catalyzed by active surface such as that provided by the sorbent. Emissions of NO_x from fluidized bed combustors are complex functions of coal composition and size, air/fuel ratio, bed temperature and pressure, gas velocity, and sorbent usage.

Three methods for minimizing NO_x emissions from fluidized bed combustors can be proposed:

- Modifying operating conditions. Decreasing excess air and temperature and increasing pressure and coal particle size can decrease NO_x emissions appreciably. Pressurized operation appears particularly effective, probably because of the increased rate of the NO reduction reactions (1-30) and (1-31) and the greater probability of CO concentrations in the vicinity of coal feed points.

- Carrying out two-stage combustion. Stoichiometric and sub-stoichiometric combustion of coal in the fluidized beds has been shown to reduce the NO_x emissions tremendously. A two-stage combustion process where the first stage is operated close to reducing condition with secondary air injection to complete combustion reduces NO_x emissions. Rearrangement of coal feed and air distribution points can probably achieve the same effect.

- Using decomposition catalysts or sorbents for NO in the fluidized beds. Argonne has injected metal oxides, Al_2O_3, ZrO_2, and Co_3O_4, into fluidized beds. Al_2O_3 and ZrO_2 did not change NO_x emissions; when Co_3O_4 was tested, NO concentration in the flue gas increased. Presumably there is a catalyst which will effect a reduction in NO_x emissions. Catalyst poisoning by sulfur oxides in the bed may be a serious problem. Some sorbent might be found which is effective for NO_x as limestone and dolomite are for SO_2. However, most nitrates, nitrites, and nitrides are not stable at the operating conditions of the fluidized bed.

Combustion Efficiency

To be able to utilize fully the high heat transfer rate in a fluidized bed and to take advantage of the air pollution control capability of fluidized bed boilers for future power generation, the combustion efficiency of the fluidized bed boilers should be at least as attractive as that of conventional p.f. boilers. Experimental evidence shows that the combustion of coal in a fluidized bed is a complex process which depends on many operational and design variables. In addition to the usual boiler losses (such as sensible heat loss from the flue gas and moisture in the fuel), the major combustion losses from a fluidized bed combustor are elutriated solid carbon and unburned combustibles such as carbon monoxide (CO) and hydrocarbons. To optimize the combustion efficiency, a careful selection of operating conditions -- bed temperature, excess air, fluidizing velocity, size distribution of fuel -- and design variables -- freeboard height, fuel feed point separation -- is essential.

The unburned carbon loss from fluidized bed combustion occurs entirely through elutriation. Thus, the amount of the carbon loss may be expected to be inversely proportional to the percentage of excess air, bed temperature, and the residence time of the coal particles in the bed, which in turn depends on the fluidizing velocity. In most cases the volatile content of the carry-over carbon is about 1.5% to 5.5% [42,55]. Consequently, the elutriated unburned carbon can be treated as solid carbon with heating value of ∿14550 Btu/lb.

When the coal is screw fed or the fuel feed points are widely spaced, experimental evidence shows variation in the flue gas compositions over the cross-sectional area of the fluidized bed. The CO concentration is usually higher over the coal feed point than at any other location, and combustion of combustibles above the bed is observed by the fact that the flue gas temperature measured five to eight inches above the bed surface shows a temperature up to 200°C higher than the average bed temperature [54]. Measurements also indicate that combustion of CO and carbon occurs up to two feet above the bed [17], and much of the CO in the gases leaving the bed is formed by reduction of CO_2 by fine carbon particles. Thus, the interpretation of data for CO loss will depend on the distance above the bed at which the gas samples were taken. Methane (CH_4) did not appear in the gas significantly until the excess air was below 5% [75,101]. The existence of other higher hydrocarbons is probably negligible because they burn more readily than methane.

Aside from the cross-sectional variations caused by the maldistribution of fuel and air, the losses from elutriated solid carbon and unburned combustibles generally relate to the other independent operational and design variables as described below. The quantitative results and the qualitative effects are summarized in Tables 1-2 and 1-3.

Fines Recycle

Ash withdrawn from the bed contains < 0.1% carbon [78,70]. Thus, most of the carbon loss occurs by elutriation of fine particles from the bed. Depending on the operating conditions and the particle size distribution of the fuel, the carbon carry-over from the bed without recycle can be very large [70,77,101], as shown in Figures 1-36 and 1-37. The type of bed material does not seem to affect carbon loss. Recycle of the fine particles back to a separate bed running at a higher temperature and higher excess air (as done by PER), or to the original bed (as done by NCB and the U. S. Bureau of Mines), are the most effective ways to reduce the carbon loss.

The experimental data obtained in a separate carbon burn-up cell for the carbon burn-up efficiency of fly ash containing 23% to 65% carbon have been correlated by regressional analysis with operating variables such as bed temperature, air rate, static bed depth, carbon rate, and inert rate [27]. The results indicate that combustion efficiency is improved by increasing bed temperature, air rate, and bed depth, and by decreasing fuel feed. The selected experimental data are plotted in Figure 1-38, and the curves shown in Figure 1-38 are reasonable approximations for design purposes. Integrated operation of a combustor and a carbon

TABLE . .

SUMMARY OF THE RESULTS OF COMBUSTION LOSSES FROM EXPERIMENTS
CONDUCTED BY OAP CONTRACTORS AND NCB LABORATORIES AT
ATMOSPHERIC PRESSURE

ORGANIZATION	REACTOR SIZE	BED MATERIAL	PARTICLE SIZE	BED TEMPERATURE (OF)	EXCESS AIR (%)	FLUIDIZING VELOCITY (FT/SEC)	FINES RECYCLE	TUBE IN BED	CARBON LOSS (%)	CO LOSS	HYDROCARBON LOSS	REFERENCE
PER	12" x 16"	Limestone BCR 1359	Bed:-10+20 mesh Coal:-1/4 x 0	1510-1630	5, 15 & 25	8-12	No	No	10-14	0.4-6.5% at 5% excess air; negligible at excess air >10%	<0.1%	[77]
PER-Carbon Burn-up Cell Tests (Burn up Cell by itself)	12" x 16"	Coal Ash	600 - 1400	1750-2000	10-65	10-15	No	No	Combustion efficiency ranges from 65% to 90%			[27]
PER-Integrated FBM/CBC Tests	FBM 20" x 6' CBC 20" x 20"	Coal Ash	Bed:-8+16 mesh Coal: -1/4 x 0	FBM: 1590-1700 CBC: 1800-2050	5-27 overall	--	Recycle to CBC	No	1.6-6.4	--	FBM: 0-370 ppm CBC: 0-600 ppm	[76]
USBM	18" diameter	refractory	Bed:-16+48 mesh Coal: 1/8" x 0	1450-1650	8.6-57.0	2.4-3.4	Recycle to original bed	Yes	<1.3	Carbon burn-up efficiency ranges from 86% to 99.8%		[15]
ESSO	3" diamet	limestone BCR 1359	Bed: 450μ, 930μ Coal: 200μ	1500-1800	5-30	2.6	No	No	3-13	450-750 ppm	--	[31]
Argonne	6" diameter	alumina	Bed: 30 mesh Coal: -14 mesh	1600	10-20	3	No	No	3-7	--	--	[3]
BCURA	12" diameter	refractory	Bed:-1/8" x 30 mesh; Coal: -1/4"+ 1/16"; 1/8" -0	1380-1800	0-90	7.2-10.6	No	Water cooling jackets	~20	--	--	[54]
BCURA	27" diameter	refractory or ash	Bed:-1/4"+ 30 mesh; Coal:-1/4" x 0; -1/8" x 0	1300-1830	0-90	7-14	Partial recycle to original bed	Yes	4-14	0.4% at 6% excess air 0.1% at 90% excess air	--	[42,55]
NCB-CRE	6" diameter	ash	Bed:-10 mesh Coal: -10 mesh	1350-1470	0-55	1-3	With & without recycle back to original bed	cooling coil	≤7% without recycle ≤1% with recycle	0.2-0.9% at 1300° F <0.3% at 1470° F	--	[7,8,18]
NCB-CRE	12" diameter	ash	Coal:-1/16"	1560	4-6	3	Partial recycle	Yes	6-7%	0.07-0.09 in flue gas	--	[71]
NCB-CRE	3 ft square	ash	Coal:-10 mesh	1300-1560	+30 to -30	2-4	No	No	10-25	<0.2	--	[70,101]
NCB-BCURA (a)	48" x 24"	ash	Bed: -1/16" Coal: -1/16"	1470	10-35	2	Recycle to original bed	Yes	1.3	<0.05%	--	[69]

NOTE: PER - Pope, Evans, and Robbins, Consulting Engineers, Alexandria, Virginia
USBM - U.S. Bureau of Mines at Morgantown, West Virginia
ESSO - ESSO Research and Engineering Company, Linden, New Jersey
Argonne = Argonne National Laboratory
BCURA - British Coal Utilization Research Association, England
NCB-CRE = National Coal Board Research Establishment, England

(a) Operating pressure at 3.5 atm.

TABLE 1-3 EFFECT OF PRIMARY VARIABLES ON COMBUSTION EFFICIENCY AND THE MECHANISMS INVOLVED

VARIABLE	CHANGE	LOSS IN SOLID CARBON	LOSS IN CO	MECHANISM
Fines recycle	-	Decrease	Increase (for internal recycles)	Further combustion; reduction of CO_2 by carbon fines
Excess air	Increase	Decrease	Decrease	Reaction equilibrium and kinetics
Fluidizing velocity	Increase	Increase	Negligible Effect	Elutriation
Bed temp.	-	Negligible effect at 1300°F - 1550°F	No effect at all in large unit; large effect at 1300°F - 1470°F in small unit	Slugging in small units; effect of fine recycles overshadow other effects; CO_2-solid carbon reaction kinetics
Freeboard height	Increase	Decrease	No effect	Entrainment and pneumatic transport characteristics
Coal feed point separation	Increase	Increase	Increase	Concentration gradient of coal in the bed

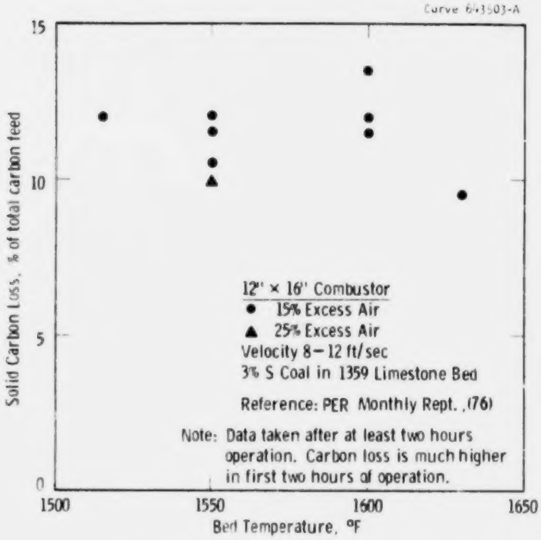


Fig. 1-36—Solid carbon loss from bed without recycle

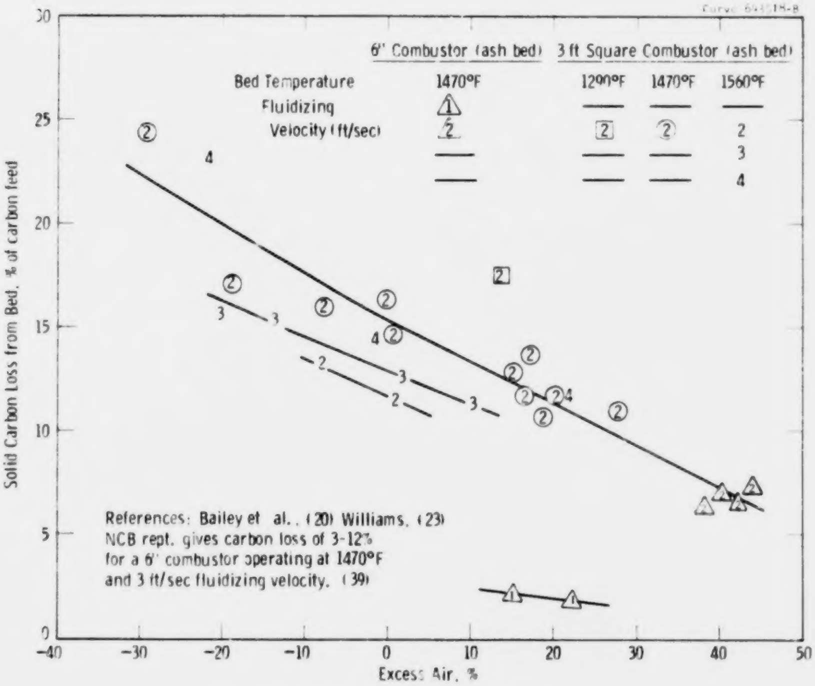


Fig. 1—37—Solid carbon loss from bed without recycle

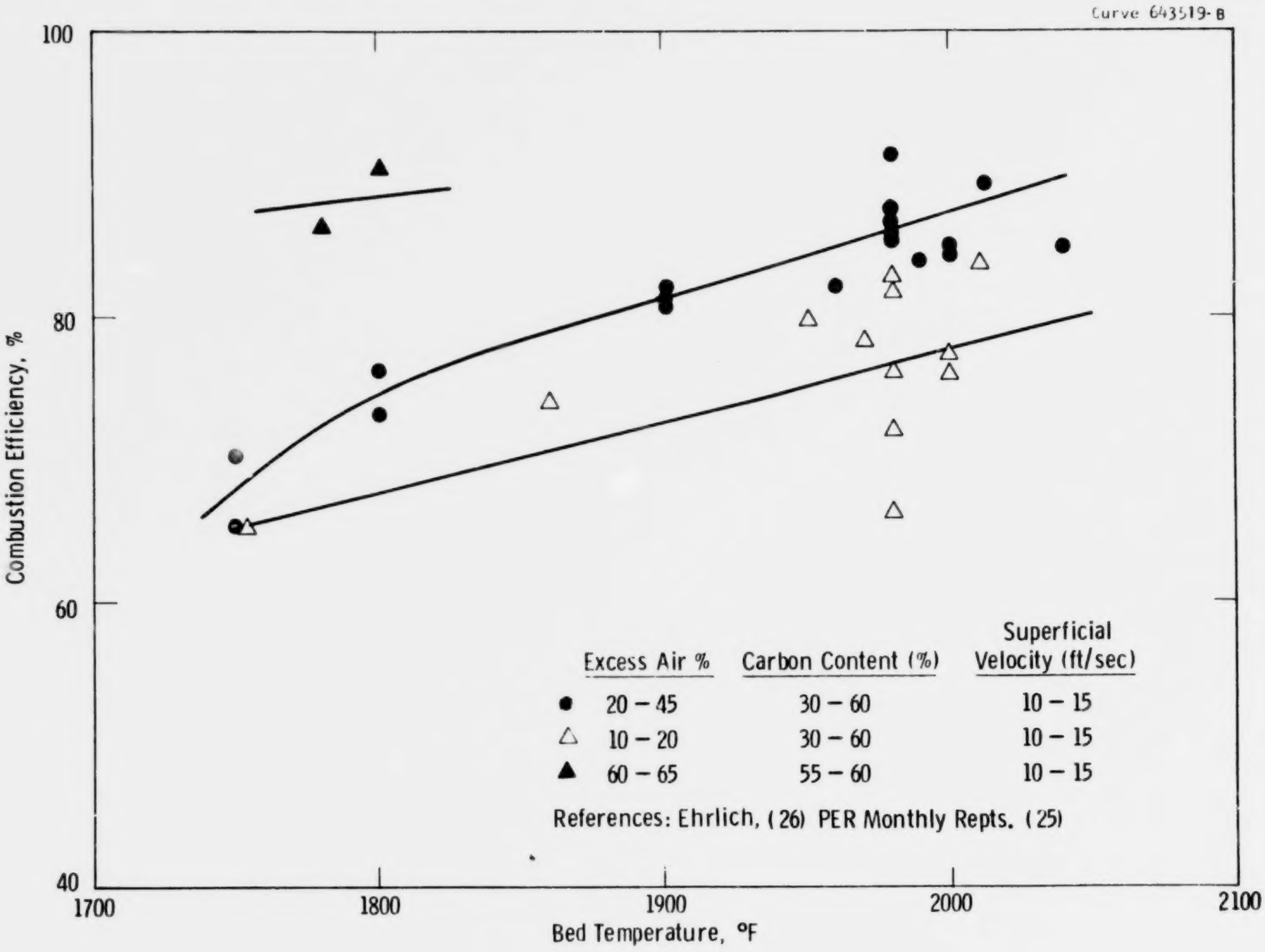


Fig. 1–38–Combustion efficiency in carbon burnup cell (ash bed)

burn-up cell was also tried by PER in their integrated FBM/CBC system [76]. The observed combustible losses, which included hydrocarbon and elutriated carbon, ranged from 1.6% to 6.4%, of which up to 10% was due to hydrocarbon loss. The high solid carbon loss was due to the short freeboard in the carbon burn-up cell necessitated by the limitations of the FBM design.

The overall solid carbon loss with internal recycle back to the same bed is shown in Figure 1-39. NCB [71] also reported a carbon loss of 6% to 7% at excess air of 6% to 4% respectively in a 12-inch diameter combustor operating at 3 ft/sec with recycle. A reduction of > 60% of carbon loss from that in Figures 1-36 and 1-37 is realizable.

Experiments performed in a six-inch combustor indicate that recycle of fines back to the original bed tends to increase CO concentration in the gas due to reduction of CO_2 by carbon fines [17]. The results from a six-inch combustor with internal recycle and a three-foot square unit without recycle are compared in Figure 1-40. There is no indication of recycle effect on the hydrocarbon losses.

Excess Air

Combustion in the fluidized bed is governed more by the excess air than by the fluidizing velocity or bed temperature in the range examined. However, appreciable combustible losses do not occur until the stoichiometric conditions have been reached. Increasing excess air tends to decrease the overall carbon content in the elutriated solids and the coal concentration in the bed as observed in a six-inch combustor [8], a 27-inch experimental rig [42], and a three-foot square combustor [101]. As a result, the overall solid carbon loss also decreases (See Figures 1-37 and 1-39).

Concentration of CO in the gas also tends to decrease with increase in excess air, as shown by experiments conducted in a six-inch combustor [7,8,18], (Figure 1-40). Up to 3% CO was observed to be present at bed surface, but most of it was burned within two feet above the bed

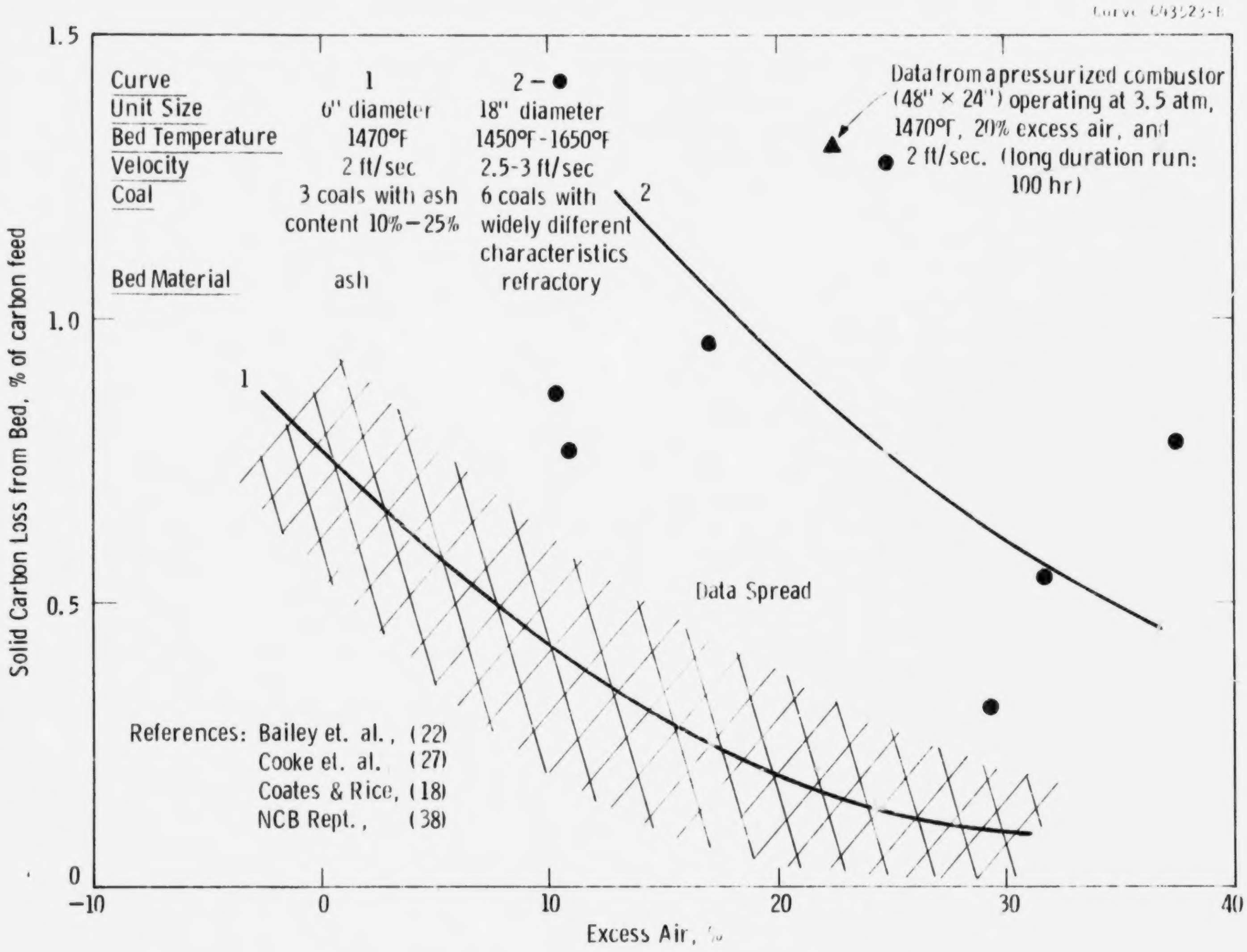
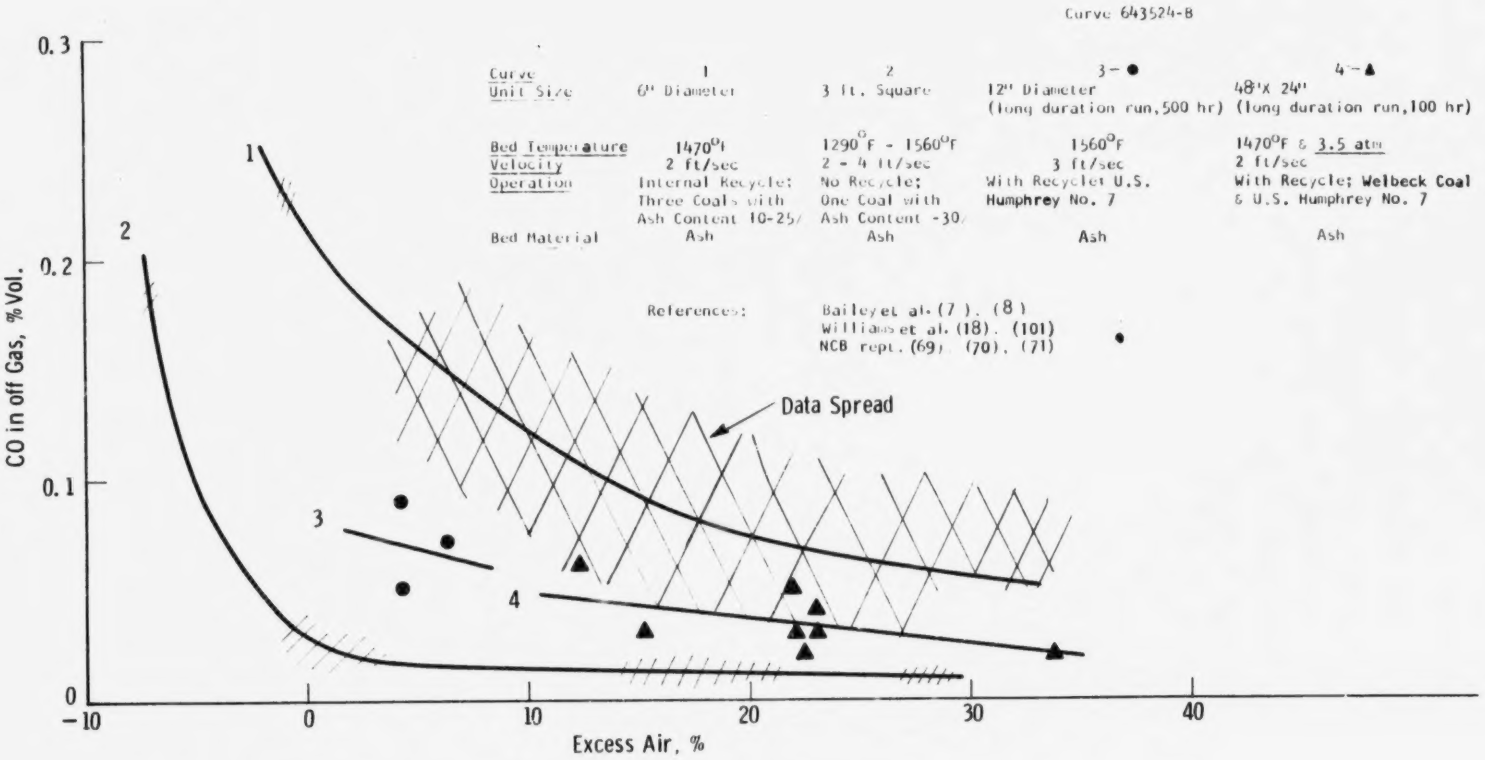


Fig. 1-39– Solid carbon loss from bed with internal recycles



Curve	1	2	3 – ●	4 – ▲
Unit Size	6" Diameter	3 ft. Square	12" Diameter (long duration run, 500 hr)	48"X 24" (long duration run, 100 hr)
Bed Temperature	1470°F	1290°F - 1560°F	1560°F	1470°F & 3.5 atm
Velocity	2 ft/sec	2 - 4 ft/sec	3 ft/sec	2 ft/sec
Operation	Internal Recycle; Three Coals with Ash Content 10-25/	No Recycle; One Coal with Ash Content -30/	With Recycle; U.S. Humphrey No. 7	With Recycle; Welbeck Coal & U.S. Humphrey No. 7
Bed Material	Ash	Ash	Ash	Ash

References:
Baileyet al. (7), (8)
Williamset al. (18), (101)
NCB rept. (69), (70), (71)

Fig. 1-40—CO loss from bed with or without internal recycles

if enough excess air was present. Measurements taken at two feet above the bed in a 27-inch diameter combustor burning 1/4 in x 0 untreated coal with partial recycle show CO concentrations of 0.4% at 6% excess air and 0.1% at 90% excess air, a value almost three times larger than that in the six-inch combustor [42]. PER [75] also reported a CO loss of 0.5% and 0.4% at 5% excess air from their experiments carried out in 12 in x 16 in and 18 in x 6 ft combustors, respectively, at 8 to 14 ft/sec without recycle; CO loss was negligible at excess air greater than 10%. However, results from the three-foot square unit measured at secondary cyclone and without recycle show much less CO loss, as shown in Figure 1-40.

The carbon and CO losses from a 48 in x 24 in bed operating under 3.5 atm pressure are also presented in Figures 1-39 and 1-40 for comparison. Operating pressure does not seem to have any significant effect on either carbon or CO losses for similar operating conditions.

Similarly, up to 0.8% CH_4 (which existed at bed surface) was completely burned within one foot of the bed surface when excess oxygen was present. The evidence of combustion taking place in the freeboard was indicated by progressively increasing concentration of CO_2 and decreasing concentration of O_2 in the freeboard.

Fluidizing Velocity

For a fixed particle size distribution of the bed material, with other operating conditions constant, increasing the fluidizing velocity tends to increase the carbon carry-over [42] (Figure 1-41), based on the terminal velocity consideration. A partial reduction of the carbon loss can thus be achieved by reducing the fluidizing velocity. The effect of velocity in a six-inch combustor [7] is also presented in Figures 1-42 and 1-43.

Although the results from the six-inch combustor show a slight dependence of flue gas CO concentration on fluidizing velocity with fines recycle, the data from the three-foot square unit operated without recycle indicate no effect of fluidizing velocity on either the concentration

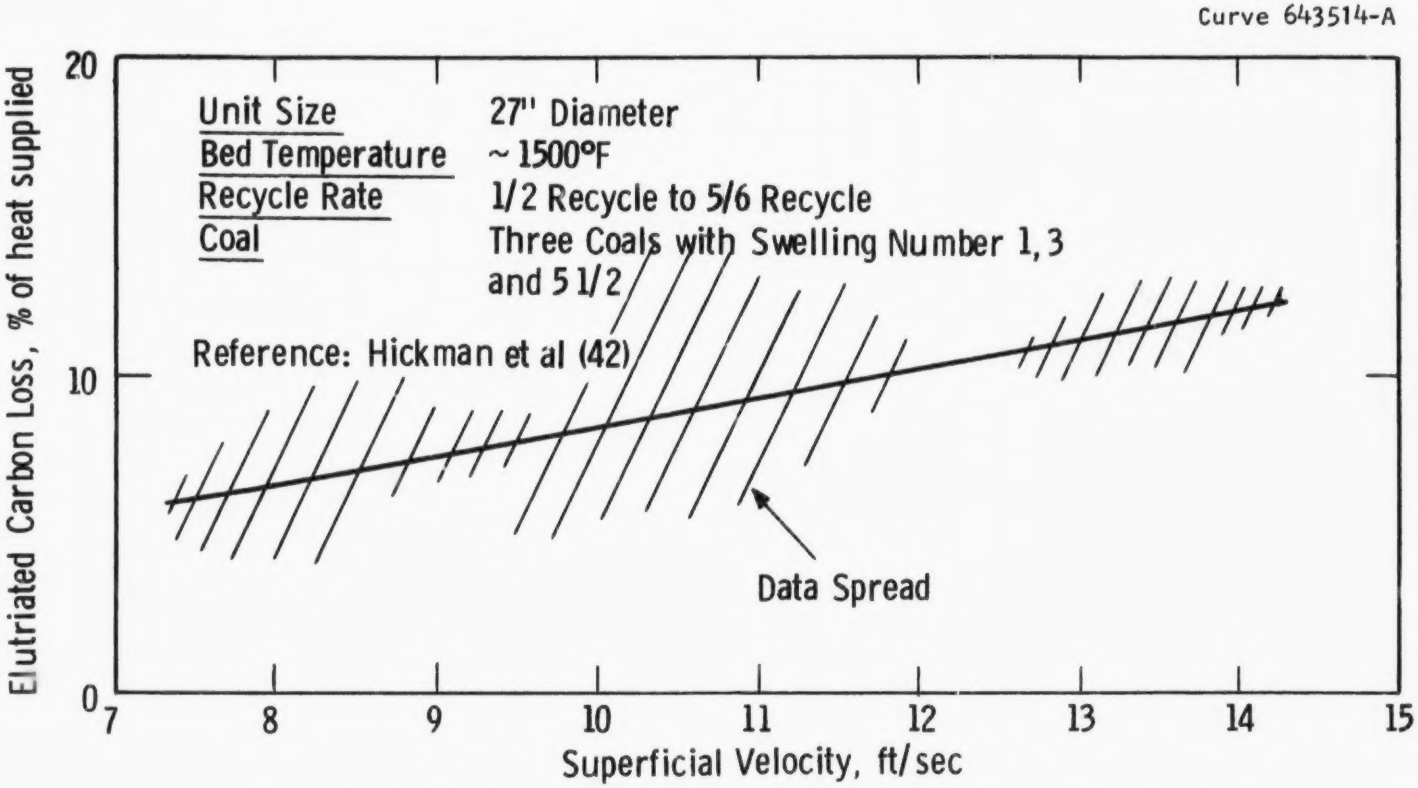


Fig. 1-41—Effect of fluidizing velocity on elutriated solid carbon loss

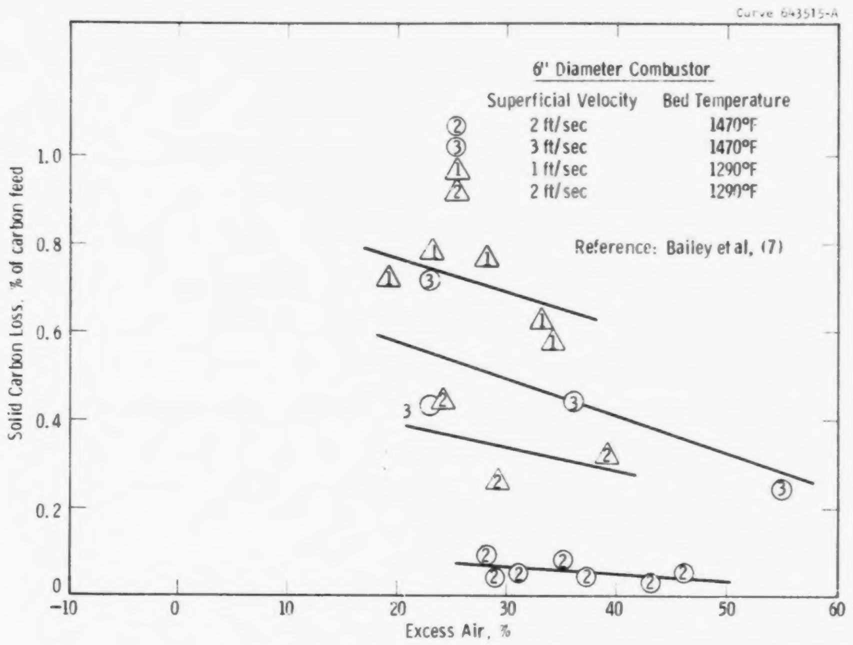


Fig. 1-42 −Solid carbon loss with fines recycles

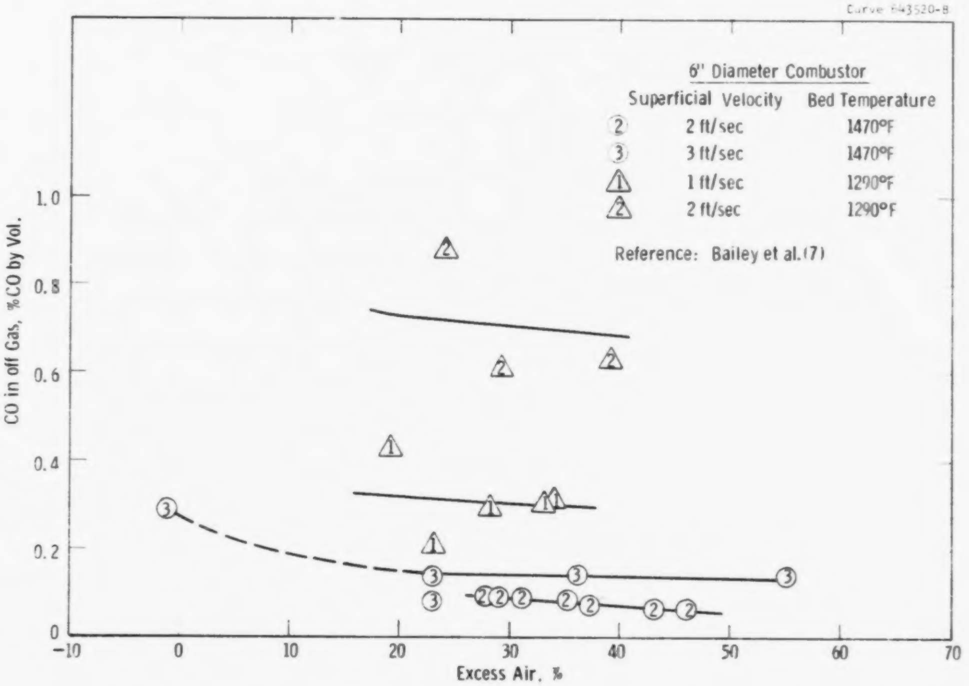


Fig. 1−43−CO loss with fines recycle

CO or hydrocarbons [101]. Here the slugging nature of the small unit may have something to do with the discrepancy; the fines recycle may prove to be a more important factor.

Bed Temperature

Within the experimental range of 1290°F to 1560°F [101], bed temperature was found to have little effect on carbon loss and no effect at all on CO concentration and hydrocarbon loss in the three-foot square unit (see Figure 1-36). However, in the smaller unit (six-inch diameter) [17], the amount of CO leaving the bed is small at bed temperatures above 1470°F and appreciable at 1290°F. Again, the discrepancy here may be due to a difference in the operation of the units: the former was operated without recycle and the latter with recycle. It may also result from differences in the kinetics of the CO_2-solid carbon reaction at these two temperature levels [17].

Particle Size of Fuel

Coal for fluidized bed combustion can be of coarse particles, as is generally the case, or of pulverized powder, as used primarily by Esso in their bench studies [31]. The combustion efficiency for these two types of coal feed will depend very much on the operating conditions. Pulverized coal particles can easily be elutriated; their rate of combustion is much higher than that of coarse particles because of their much higher surface/weight ratio. The comparison of coal combustion efficiency between these two types of operation is difficult because of a lack of common basis. However, a comparison of fluidized bed compactness can be made on the basis of the heat release rate per unit bed area. Specifically, 13% carbon loss was reported [31] for pulverized coal feed at a superficial velocity of 6 ft/sec, 1600°F bed temperature, and 10% excess air. The same amount of carbon loss was attained at about 12 ft/sec superficial velocity for the coarse coal feed of top size 1/4-in particles at the same bed temperature and excess air (Figure 1-36). Assuming that the carbon burn-up cell will be equally efficient in both cases, twice the amount of the primary fluidized bed area will be required for the pulverized unit to achieve the same overall heat release rate.

Freeboard Height

In comparing the results of solid carbon loss from different experimental units, the freeboard height of each unit must be distinguished from its respective transport disengaging height (TDH). When gas bubbles erupt at the surface of the bed, solid particles are splashed far into the freeboard. If the gas exit is located too close to the bed surface, the solid particles entrained in the gas stream usually are high. Hence, it is desirable to locate the gas exit beyond the so-called TDH, where the entrainment of solid particles becomes constant and depends solely on pneumatic transport characteristics. The empirical relationship for determining TDH for vessels of different diameters operating at different velocities is given by Kunii and Levenspiel [60]. Clearly, comparing the results of solid carbon loss from different experimental units without regard to their freeboard height can lead to error. Unfortunately, however, the correlation of solid carbon loss with freeboard height and TDH has not yet been established.

Coal Feed Point Separation

It is predicted that a wide spacing between coal feed points will cause variation in coal concentration across the fluidized bed, resulting in a decrease in the combustion efficiency within the bed due to high combustible losses [44]. A mathematical model coupled with some experimental data shows no net increase in the elutriation rate for coal feed spacings up to three feet apart. For spacings greater than that, the elutriation rate of solid carbon starts to increase, and reducing conditions start to develop around the area immediately next to the coal feed points. Thus the boiler tubes will have to withstand both reducing and oxidizing conditions in the same bed.

The value of the coal diffusion coefficient is proportional to fluidizing velocity. Thus, for higher operating velocity the optimum coal feed point spacing may actually be greater than three feet. The value of the diffusion coefficient is also smaller when tubes are inserted in the bed. The optimum coal feed point spacing will therefore have to be determined individually for each boiler at specific operating conditions and with specific designs. Without actual experimental data, 10 ft^2 bed area per coal feed point is recommended as a reasonable and conservative choice [44].

Other Secondary Variables

- Concentration of coal in the bed. Increasing the concentration of coal in the bed or increasing the coal feeding rate tends to increase the elutriation rate of solid carbon and decrease the combustion efficiency.
- Bed height. Increasing the bed height will increase the combustion efficiency by decreasing solid carbon loss and CO loss because of increased residence time of coal particles in the bed.
- Heat transfer surface arrangement above the bed. Too much heat transfer surface immediately above the bed, to serve as a splash screen, tends to quench the flue gas too fast and prevent further combustion of fines and CO above the bed. This will result in higher carbon and CO losses. Careful design in this respect is necessary, and more experimental data are required to provide adequate design information.
- Unit size. Experimental data obtained from smaller units tend to give higher carbon loss due to slugging, while the coal concentration gradient in the bed, which is important in larger units, is relatively unimportant in smaller units. The manifestation of the experimental data must be carefully analyzed for design purposes.

Summary

Combustion efficiency for fluidized bed combustion boilers depends on operating conditions, e.g., bed temperature, excess air, fluidizing velocity, size distribution of fuel; and design variables, e.g., bed depth, freeboard height, fuel feed point separation. The major combustion losses from a fluidized bed combustor are elutriated solid carbon and unburned combustibles, such as carbon monoxide (CO) and hydrocarbons. Recycle of fines back to the original bed or to a separate carbon burn-up cell operating at higher temperature and excess air is essential for efficient operation of fluidized bed combustors.

Carbon monoxide and hydrocarbon losses are most sensitive to the excess air used for combustion. However, appreciable combustible losses do not occur until the stoichiometric conditions have been reached. Solid carbon loss occurs primarily by elutriation of fine particles from the bed, and this is most sensitive to changes in fluidizing velocity. Presumably the selection of particle size and its distribution will also be important. Operating pressure and temperature do not seem to have a pronounced effect on carbon and CO losses within the ranges tested.

Design parameters such as freeboard height and coal feed point separation are also important in improving overall combustion efficiency. Unfortunately, few data are available, especially for cases where tubes are inserted in and above the bed.

Most of the studies concentrated on coal combustion in small units at atmospheric pressure with low velocities and shallow beds. Although the PER experiments were run at higher velocities (12 to 14 ft/sec), there were no immersed tubes in the bed. Extrapolation of available data for design purposes is possible but risky. More data collected under pressurized operation with high velocities (up to 15 ft/sec) and deep beds (up to 15 ft) are needed.

The results of experiments conducted by OAP contractors and NCB laboratories were summarized in Table 1-2. The effect of changing primary variables on combustion efficiency and the mechanisms involved were presented in Table 1-3. The assumptions on combustible losses, derived from evaluation of the available data presented in Table 1-2, for designing industrial, atmospheric pressure, and pressurized fluidized bed boilers are presented in Table 1-4.

TABLE 1-4

COMBUSTIBLE LOSSES ASSUMPTIONS FOR DESIGN OF THE INDUSTRIAL, ATMOSPHERIC PRESSURE, AND PRESSURIZED FLUIDIZED BED BOILERS

	INDUSTRIAL BOILER	ATMOSPHERIC PRESSURE UTILITY BOILER	PRESSURIZED UTILITY BOILER
SOLID CARBON LOSS			
From primary bed	13%	13%	6%
From CBC	90% carbon burn-up efficiency	85% carbon burn-up efficiency	90% carbon burn-up efficiency
CO LOSS	Included in carbon loss	Included in carbon loss	<0.5% in flue gas
HYDROCARBON LOSS	Negligible	Negligible	Negligible
COAL FEED POINT SEPARATION	10 ft^2 per feed point	10 ft^2 per feed point	10 ft^2 per feed point

Heat Transfer

One of the major advantages of fluidized bed boilers is the high overall heat transfer coefficient between the bed and the immersed heat transfer surfaces, which permits a boiler design of higher volumetric heat release rates. A heat release rate on the order of 500,000 Btu/hr/ft^3 furnace volume or higher is possible in fluidized bed boilers, compared to ∿20,000 Btu/hr/ft^3 in conventional boilers. Consequently, a more economical boiler with smaller boiler size and less tube surface can be constructed using fluidized bed combustion technology. Pressurized operation, where deeper beds are feasible, offers even further size reduction and economic incentives.

The literature contains a wealth of data on heat transfer between fluidized beds and immersed surfaces. Valentik and Williams [99], of NCB, have reviewed the recent experimental data from the literature and from the work done by NCB laboratories. The Massachusetts Institute of Technology (MIT), under contract to OAP, has compiled published investigations with a listing and discussion of the various theoretical, empirical, and semi-empirical expressions for predicting heat transfer coefficients in fluidized beds [13]. Numerous other articles and books review heat transfer in fluidized beds. This discussion is a review of the experimental studies performed by OAP contractors and by NCB laboratories in England which are directly related to fluidized bed combustion. Wherever adequate data from other sources are available, they are also presented for comparison.

A reasonable assumption of the heat transfer coefficient is essential for design purposes. However, the reported heat transfer coefficients range from as low as 30 Btu/ft^2/hr/°F to as high as 170 Btu/ft^2/hr/°F, because of varying operating conditions and design variables. In the case of high-temperature combustion in fluidized beds where the radiant contribution is important, the observed heat transfer coefficients can be several times higher. Therefore, our primary objective here is to evaluate the effects of operating variables and design parameters on the overall heat transfer coefficient in fluidized beds so that the knowledge can be utilized in designing fluidized bed boilers for power generation and air pollution abatement.

Two principal resistances in series must be considered in evaluating the overall heat transfer coefficient: bed-tube heat transfer resistance and tube-water heat transfer resistance. For nucleic boiling on the water side, heat transfer coefficients on the order of 1000 Btu/ft^2/hr/°F may be expected. If the water-side or steam-side conditions are forced circulation, as may be expected in economizer or superheater and reheater designs, the water-side coefficients are usually less (300-600 Btu/ft^2/hr/°F) but can be reasonably estimated from the existing empirical relations, such as the Dittus-Boelter equation [43]. The bed-tube heat transfer coefficient, however, depends to a great extent on the operating conditions (bed temperature, fluidizing velocity, bed particle size, solid circulation patterns in the bed); on the design parameters (bed size, tube pitch/diameter ratio, number of rows of tubes, triangular or square pitch, vertical or horizontal tube); and on the properties of reactants (thermal conductivity, thermal diffusivity, density, and heat capacity of the bed material and gas in the bed). The operating and design variables are classified into primary and secondary variables, depending on their relative importance in influencing the heat transfer coefficients.

Primary Variables - Operating Variables

Bed Temperature. Changing the bed temperature affects the overall heat transfer by altering:

- The thermal conductivity of gas and thermal diffusivity of the bed material
- The radiation contribution portion of the total heat transferred
- The overall temperature difference between the bed and the immersed heat transfer surfaces.

The heat transfer between the bed and the immersed tubes constitutes two portions, the convective and the radiant heat transfer. The magnitude of convective heat transfer depends on the particle size distribution, residence time of the particles at the heat transfer surface, the density of particles in the bed, and the physical properties of the gas medium. Botterill [12] showed that particle-to-surface contact areas are too small for significant heat transfer to occur by conduction through points of solid contact. Experimental results show that the heat transfer takes place primarily by conduction through the thin gas film between the surface and the particles. In other cases, the residence time of the particles near the heat transfer surfaces can be a limiting factor. Hence, the heat transfer coefficients were found to be higher when gases with higher thermal conductivities [13] and solids with higher heat capacities were used at given experimental conditions. Experimental data indicate that the heat transfer coefficient, h, depends on the thermal conductivity of the gas, k_g, and the heat capacity of the solid, C_s, in the following ways:

$$h \alpha k_g^{0.33} \text{ to } k_g^{1.0}; \; h \alpha C_s^{0.25} \text{ to } C_s^{0.8}.$$

Experimental measurements at bed temperatures between 1000°F and 1800°F suggest that radiation in a fluidized bed may be treated as a black body and is additive to the convective heat transfer [42]. At ∿1500°F bed temperature, the radiation contribution ranges from 10% to 50% of total heat transferred in the bed, depending on the sizes of bed particles and the fluidizing velocity.

Hence, the heat transfer coefficient increases with an increase in bed temperature because of the increase in thermal conductivity of the gas, the increase in thermal diffusivity of the particles, and the increased contribution of radiation. The quantitative effect of the bed temperature on the heat transfer coefficient, based on the studies by Kharchenko and Makhorin [56] and Peel [73], is presented in Figure 1-44. The bed temperature effect is complicated and depends on additional

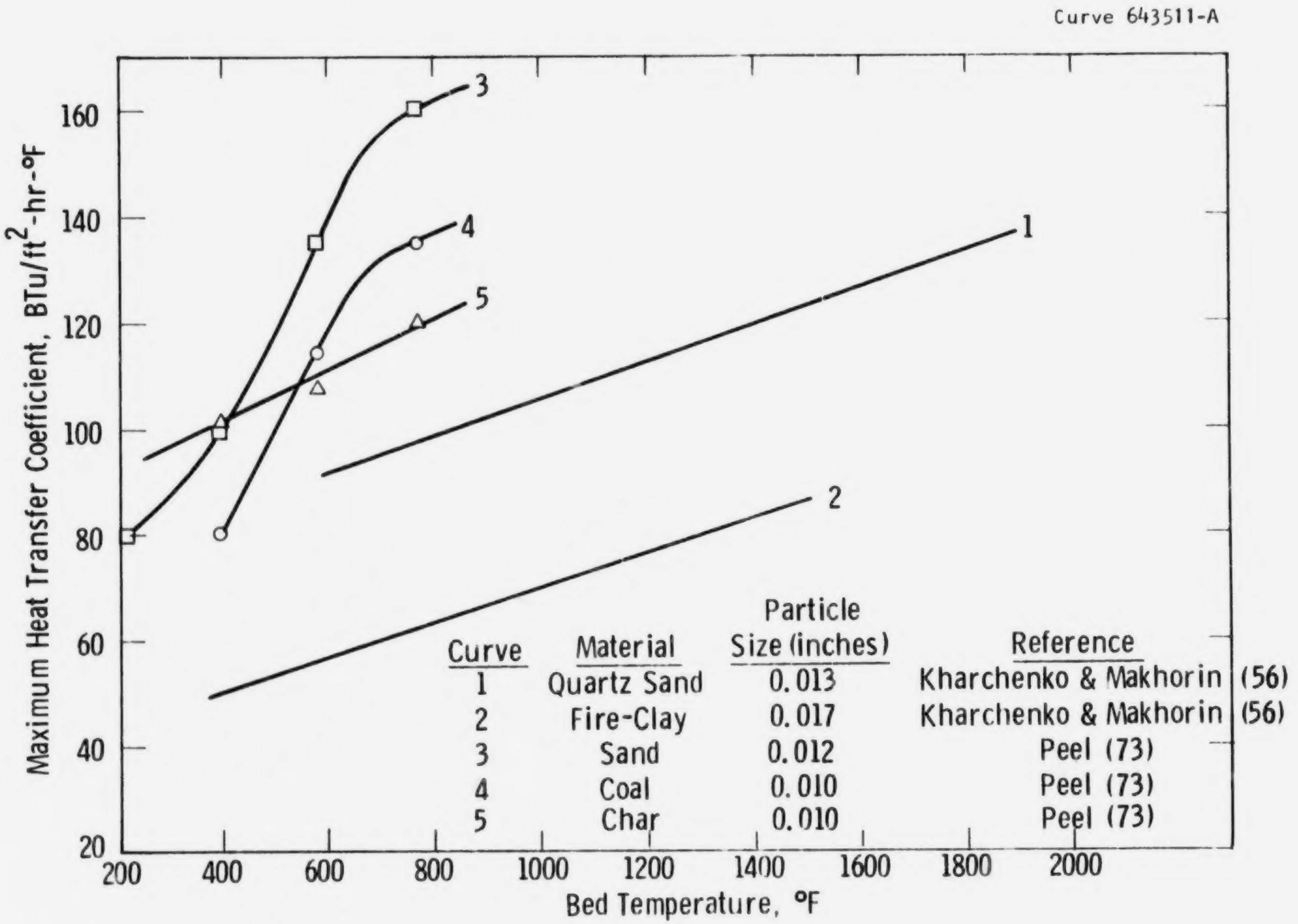


Fig. 1-44—Effect of bed temperature on heat transfer coefficient

variables such as bed material and particle size. Note that the data are not obtained in a fluidized bed combustor. In fact, few data from fluidized bed coal combustion which elucidate the bed temperature effect are available. Fortunately, a comprehensive study of bed temperature effect is not necessary for the fluidized bed combustion of coal because the temperature range of interest is quite narrow (1400°F to 1700°F in the primary combustor and 1800°F to 2000°F in the carbon burn-up cell) and can probably be accounted for satisfactorily by a radiation contribution.

Operating Pressure. Running at higher pressure will provide better fluidization and heat transfer because the density of the gas is higher. Because of the scarcity of experimental data, the quantitative improvement is not clear.

Particle Size. Experimental evidence [54,56,105] indicates that decreasing the particle diameter causes an increase in the heat transfer coefficient due to an increase in particle mobility and reduction in the thickness of the gas film between the immersed heat transfer surfaces and the aggregates of particles. The particle size distribution will also affect the magnitude of the heat transfer coefficient: a wide size particle range tends to give better fluidization and higher particle mobility and agitation.

The reported dependence on average particle diameter ranges from $d_p^{-0.17}$ to $d_p^{-0.96}$ [13,43]. The effect of average particle size is shown in Figure 1-45 according to Wright et al., [102]. The data are for the convective heat transfer component only; the radiation contribution is subtracted from the total heat transfer by assuming the fluidized bed to be a black body at its bed temperature [42]. The data show that the convective heat transfer coefficients start to level off at a mean particle size larger than ∿0.07 in. The design particle size in the fluidized bed combustor is -1/4 in with a mean particle size around 0.07 in (≈1800μ). Variation of the heat transfer coefficient due to changes of mean particle size from attrition, elutriation, and screening operations is thus estimated to be less than 10%. At smaller mean particle sizes, the effect may be more pronounced.

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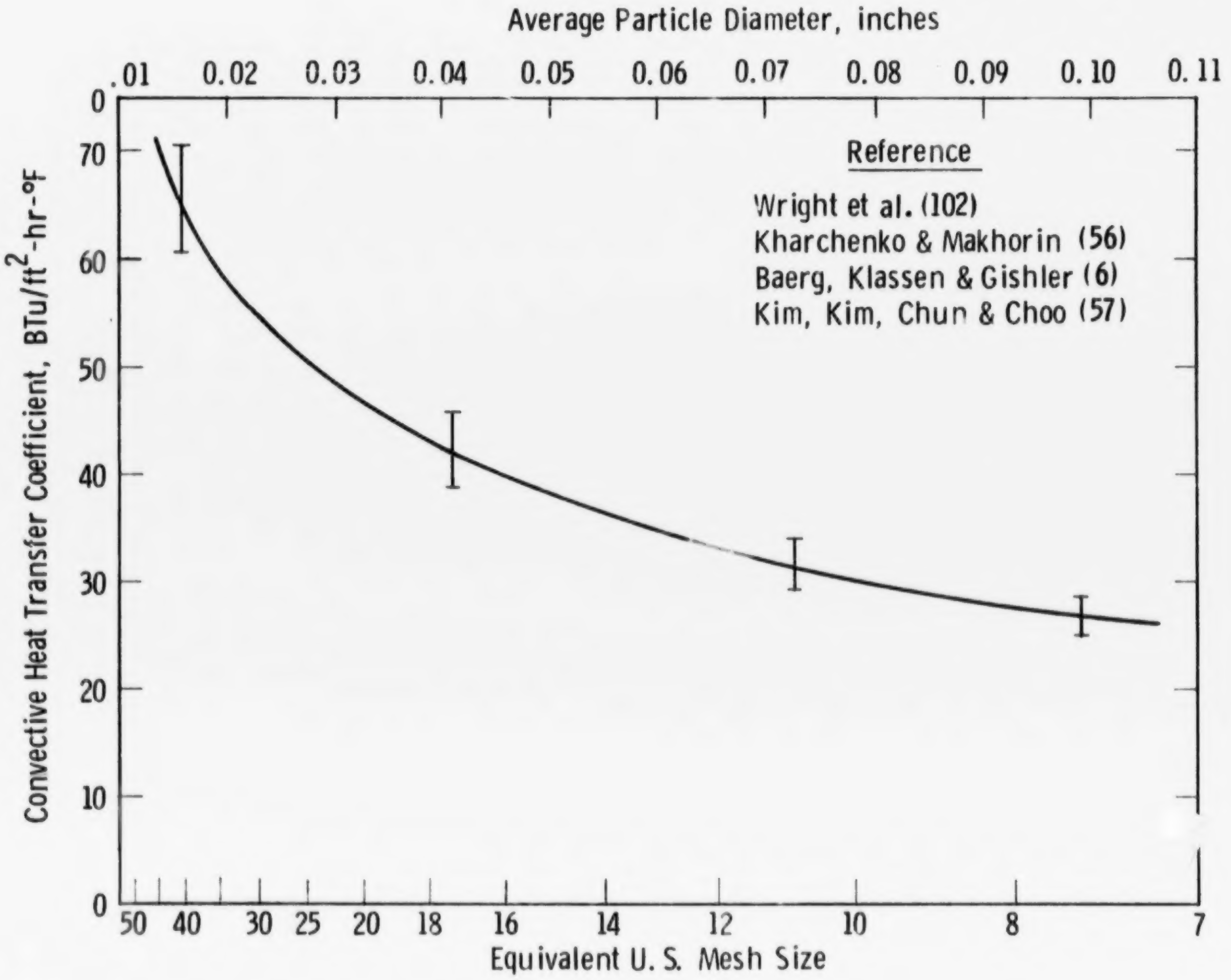

Fig. 1-45--Effect of particle diameter on heat transfer coefficient

Fluidizing Velocity. As the fluidizing velocity is increased over the minimum fluidizing velocity, the heat transfer coefficient first increases rapidly due to increased agitation in the bed until a maximum is reached at about two to nine times the minimum fluidizing velocity [56,58,74], depending on the particle size in the bed. Then the heat transfer coefficient remains essentially constant for a short range. Further increases in fluidizing velocity cause a decrease in the heat transfer coefficient, because the decrease in particle concentration in the bed overtakes the effect of increasing particle agitation. In a bed of larger particle size, where the large particles tend to segregate out at the bottom of the bed and become sluggishly fluidized at lower fluidizing velocities, the effect of superficial velocity is especially clear. Figure 1-46 shows the experimental measurements [42] of heat transfer to the containing walls in a fluidized bed burning an untreated coal of 1/4 in x 0 size at different bed depths. In the lower part of the bed, where a sluggishly fluidized bed of bigger particles is observed, the heat transfer coefficient is lower and highly dependent on the fluidizing velocity for both horizontal and vertical tubes in the bed. In the upper part of the bed, where the quality of fluidization is good, higher heat transfer coefficients are obtained with less dependence on velocity. In general, stronger dependence on velocity was observed with vertical tubes in the bed. For coarser particles and higher bed temperatures, the region for a constant heat transfer coefficient can cover quite a wide range of gas stream velocities at fully fluidized conditions.

The rate of heat transfer to the immersed surfaces was found to be generally higher than that to the walls and was also dependent on fluidizing velocity, as shown in Figure 1-47 where only the measurements at the fully fluidized section were shown. The operating conditions and the heat transfer coefficients obtained are summarized in Table 1-5.

It is important that, in order to avoid segregation of particles in the bed, the design fluidizing velocity must be at least larger than the minimum fluidizing velocity of the biggest particle in the bed. For example, the minimum fluidizing velocity of 1/4-in particles at the

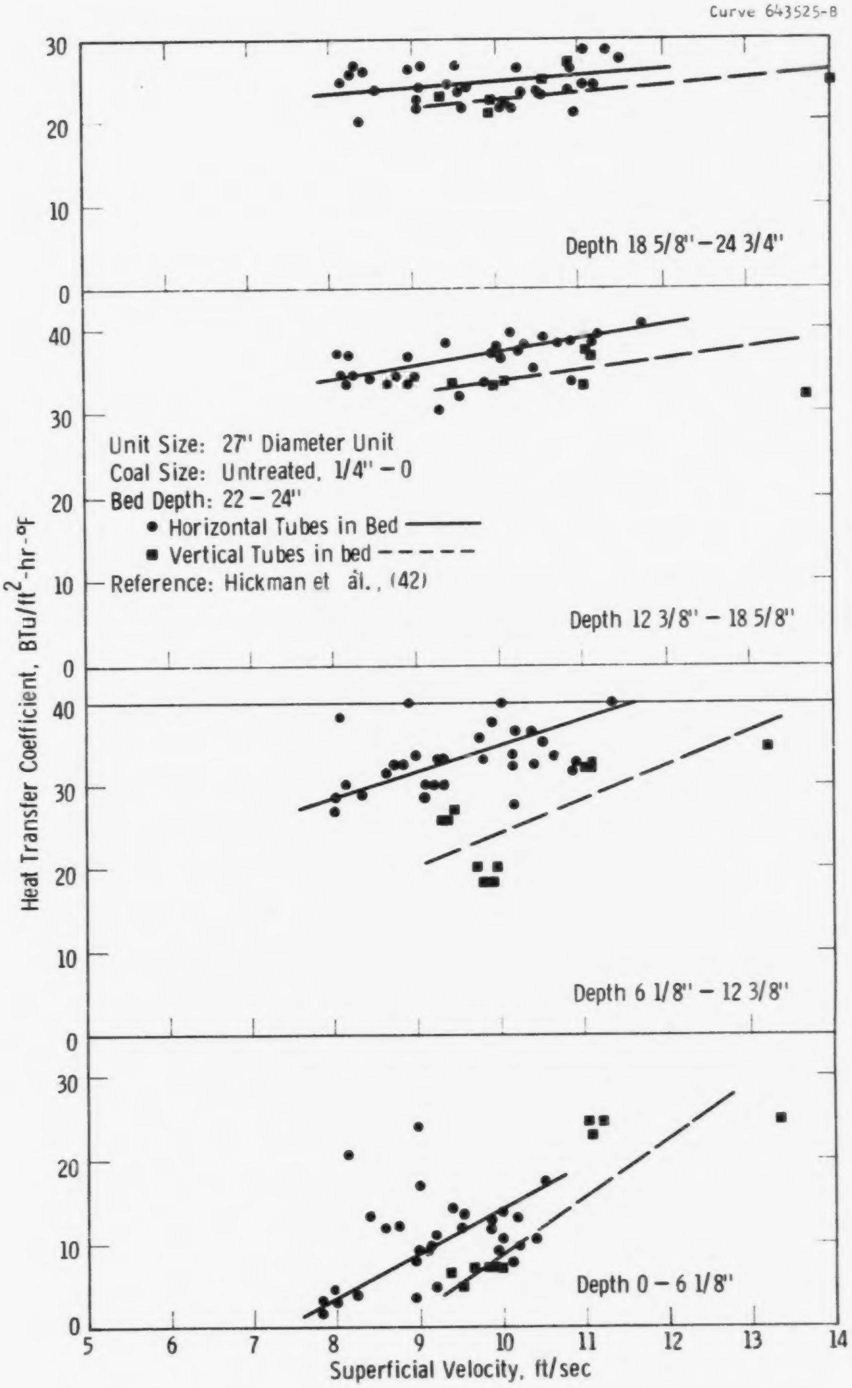

Fig. 1-46—Effect of fluidizing velocity on heat transfer coefficient to walls

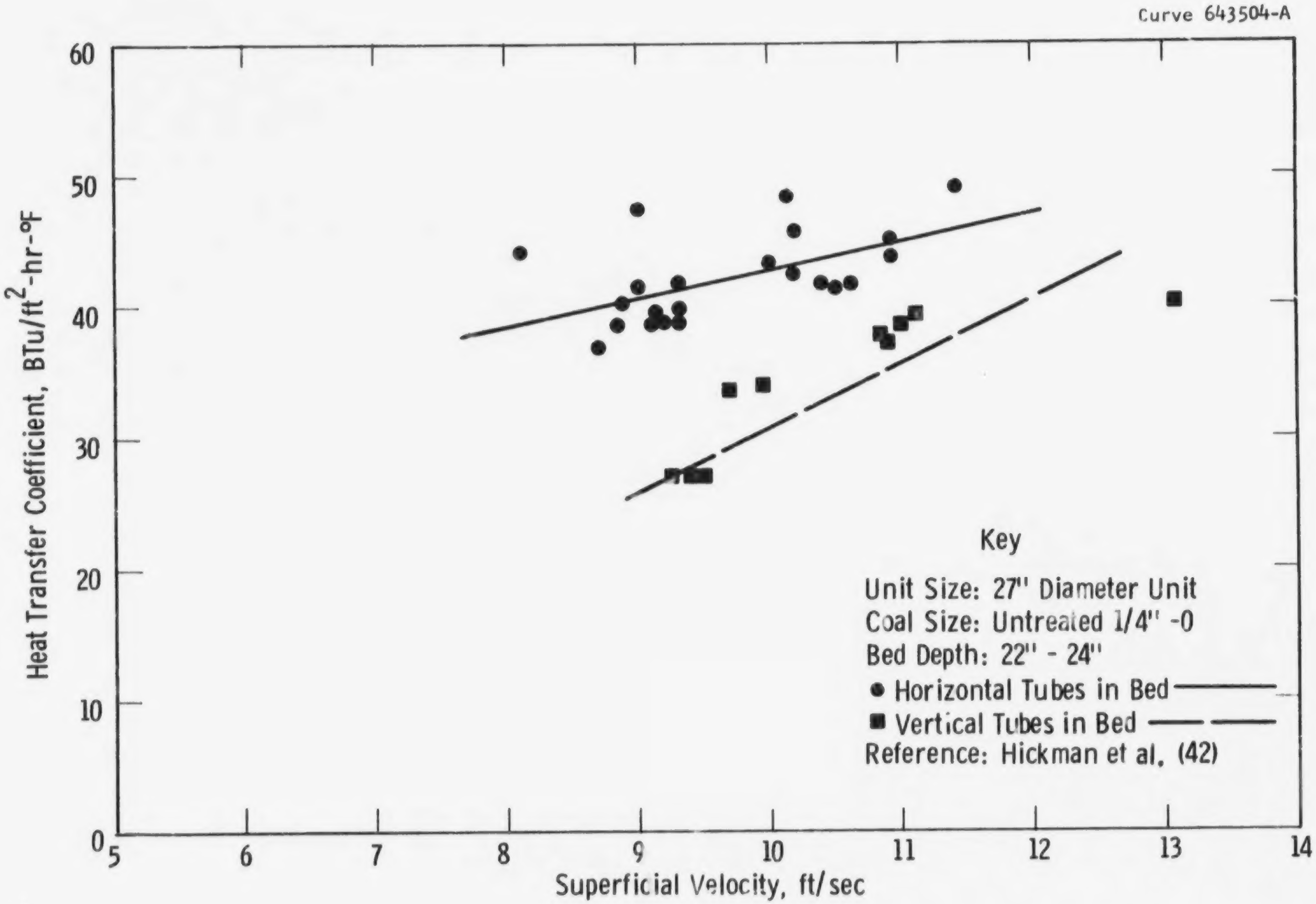

Fig. 1-47 —Effect of fluidizing velocity on heat transfer coefficient to immersed surfaces

TABLE 1-5

SUMMARY OF HEAT TRANSFER MEASUREMENT IN FLUIDIZED BEDS AT
ATMOSPHERIC PRESSURE PERFORMED BY OAP CONTRACTORS AND NCB LABORATORIES

ORGANIZATION	REACTOR SIZE	BED TEMP. (°F)	BED MATERIAL	PARTICLE SIZE	FLUIDIZING VELOCITY (FT/SEC)	TUBE IN BED	IN BED	AT WALLS	ABOVE BED	REFERENCE
BCURA	27" diam.	700°C-1090°C	Refractory or ash	Inert: - 1/4" + 30 mesh coal: - 1/4" - 0	6-14	Vertical horizontal	25-40 40-50	35 35-40	---	[42]
BCURA	12" diam.	750°C-982°C	Refractory	Inert: 18-30 mesh coal: - 1/8" - 0	7-11	Horizontal	40-57	20-30	---	[54]
NCB-CRE	3 ft. sq.	Room temp.	Coal ash	- 10 mesh	1-4	Horizontal	20-50	---	30% less	[43]
NCB-CRE	12" square	800°C-850°C	Coal ash	- 10 mesh	2-2.5	Horizontal tube & loop	78-85	---	---	[85]
USBM	18" diam.	1450-1650°F	Al_2O_3	- 16 + 48 mesh	2.4-3.4	Horizontal	54-67	---	Avg. 53	[15]
PER	12" x 16"	1450-1650°F	Coal ash or limestone BCR 1359	Ash: -7+ 14 mesh limestone: - 10+20 mesh	12-14	Bayonet heat exchanger	47	---	---	[75]

NOTE: BCURA = British Coal Utilization Research Association, England.
NCB-CRE = National Coal Board - Coal Research Establishment, England.
USBM = U.S. Bureau of Mines at Morgantown, West Virginia.
PER = Pope, Evans, & Robbins, Consulting Engineers, Virginia.

experimental conditions is ~10 ft/sec; the fluidized bed with the same particle size distribution and operating at a velocity > 10 ft/sec will have better particle mixing and consequently a more uniform heat transfer coefficient in the bed (Figures 1-46 and 1-47). Segregation of bed particles will result in a lower overall heat transfer coefficient, temperature gradient in the bed, and possible agglomeration due to sluggish agitation. However, this is not a totally detrimental effect. Segregation of bed particles which results in lower overall heat transfer coefficient will be helpful during boiler turn-down (see Appendix E) if particle agglomeration can be prevented.

The relationship between particle segregation and fluidizing velocity is not very well understood and should be further investigated, especially in pressurized operation with deep beds, where the phenomenon is expected to be enhanced. Cold model studies in this area are being conducted to gain more understanding of the problem and to provide some information for design.

Primary Variables - Design Parameters

Position of Heat Transfer Surface. Many investigators have reported variations in the heat transfer coefficient relative to the position of heat transfer surfaces in the fluidized bed, particularly the differences in coefficients for vertical and horizontal tubes in the bed [65,99] and the differences between containing walls and immersed heat transfer surfaces [42,55,56], (Figures 1-46 and 1-47).

Morgan [65] reported an increase of heat transfer coefficient up to 40% greater for vertical tubes than for horizontal tubes, depending on the average particle size employed (Figure 1-48). This contradicts the data presented in Figures 1-46 and 1-47. Although the vertical tube arrangement usually gives worse fluidization because of a possible channeling effect, the consistently higher heat transfer coefficient for vertical tubes here is probably due to the stagnant layer of particles on the upper surface of the horizontal tubes which renders part of the

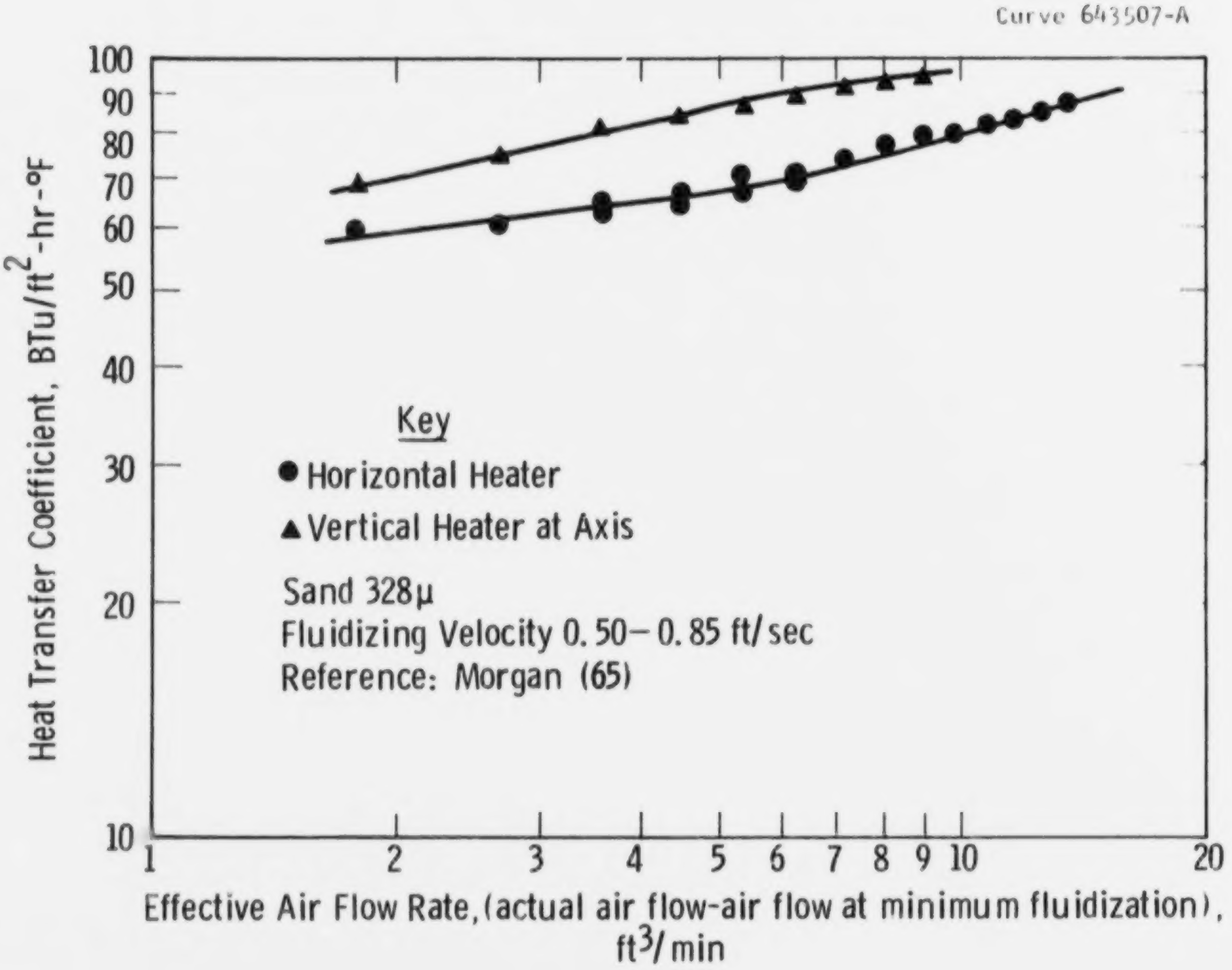

Fig. 1-48—Variation of heat transfer coefficient with position of heat transfer surface

surface inactive. Westinghouse cold model study confirms this hypothesis [53]. However, the study by BCURA burning 1/4 in x 0 untreated coal in an ash bed with vertical and horizontal tubes sized at 2-3/8 in O.D. had the opposite result (Figures 1-46 and 1-47), where the heat transfer coefficient to vertical tubes was shown to be up to 30% below that of horizontal tubes. The explanation of this apparent contradiction may be that the stagnant layer of particles on the upper surface of the horizontal tubes was absent in BCURA's studies because of the high velocities and the large particles employed. The stability of this stagnant layer depends on particle size/tube diameter ratio and the violence of agitation. Presumably, the tube pitch/diameter ratio also plays an important role here.

The difference in the heat transfer coefficients at walls and at the internal heat transfer surface, however, is caused by an entirely different mechanism. For a gas-solid fluidized bed operating at atmospheric pressure, the solid agitation is controlled almost entirely by bubbling phenomena of gas bubbles which carry the solid particles with them in the wake until they finally burst at the bed surface. In addition, the gas bubbles tend to pass from the bottom to the top of the fluidized bed through the lowest resistance pass, i.e., clear of the containing walls. To maintain the balance in the bed, particles have to move downward in the particulate phase and thus set up a mass circulating flow pattern [52] where the solid particles are swept rapidly upward in the middle of the bed by the gas bubbles and then move slowly downward near the walls. Hence, the same particles near the walls remain in close contact with the walls for a relatively long period of time without replacement and cause the heat transfer coefficient to drop (Figures 1-46 and 1-47).

Conflicting results were also reported for the heat transfer coefficient of the horizontal tube banks at the top and bottom of the fluidized bed. Highley et al. [43] found, in their studies of heat transfer in a three-foot square cold ash bed with a bed depth of three feet, that raising a bundle from the lower to the upper half of the bed increased the heat transfer coefficient 20% and also observed some segregation of particles in the bed at about two times the minimum fluidizing velocity. Using 1/4 in x 0 size particles at low fluidizing velocity, Hickman et al. [42] reported a difference of heat transfer coefficient for heat transfer surfaces located at different bed depth due to segregation of bed particles, which resulted in a sluggishly fluidized portion at the bottom of the bed (see Figure 1-46). There is, however, no other strong evidence that the heat transfer coefficient should be different for tubes at different bed depth at good fluidization conditions.

For tubes exposed immediately above the bed and still in the splashing range of the bed material, a drop of 30% in the heat transfer coefficient was noted in room-temperature ash bed experiments [43]. However, in combustion of coal at bed temperatures in the range 1450°F to 1650°F, Coates et al. [15] reported approximately the same overall heat transfer coefficient, 53 $Btu/ft^2/hr/°F$, both for the tubes immersed in the bed and for the tubes a few inches above the bed. The discrepancy may come from either the confusing definition of bed height or from the additional combustion of combustibles above the bed and the radiation effect in the "hot" experiments.

In summary, the heat transfer coefficient for the immersed surfaces in the bed is at least 10% greater than that for the immersed surfaces at walls. Efficiency in heat transfer for vertical tubes versus horizontal tubes can be quite different depending on other operating and design variables such as particle size/tube diameter ratio, superficial fluidizing velocity, and tube pitch/diameter ratio. For tubes exposed immediately above the bed and still in the splashing range of the bed material, a drop of up to 30% in the heat transfer coefficient is possible if the additional combustion of combustibles above the bed and the radiation effect are absent. More data are required to permit accurate design of heat transfer surfaces above the bed and in the freeboard.

Tube Arrangement. Tube pitch/tube diameter ratio is another important variable to be considered for design. Reducing the gap between adjacent tubes tends to reduce the mobility of the particles and hence the heat transfer coefficient. Increasing the number of the staggered rows has the same effect. In reducing the distance between adjacent tubes in stages from 11 in to 0.6 in, a reduction of 25% in the heat transfer coefficient was observed [43]. Adding an extra row to increase the staggered rows from two to three rows when the gap was 2.5 in reduced the coefficient by a further 15%. The range of tube pitch/tube diameter ratio employed is two to eight with tube sizes of 1-3/8 in O.D. and 2-3/8 in O.D. The result of the measurements at room temperature in a vessel three feet square is shown in Figure 1-49. The results can also be correlated in terms of the narrowest gap between tubes in the bundle, as shown in Figure 1-50 [64]. The heat transfer coefficient decreases continuously for about 25% as the gap is reduced from 11.1 in to 0.59 in without any sharp drop indicating a marked decrease in particle mobility. The particle size/narrowest gap (between tubes) ratio is probably important; however its effect has not yet been studied.

The extent of the effect of using square pitch rather than triangular pitch is not clear.

It is advantageous to keep the pitch/diameter ratio as small as possible. For example, to lower the pitch/diameter ratio from eight to two, an increase of 16 times of heat transfer surface per unit volume of bed,decreases the heat transfer coefficient by only 18% (Figure 1-49). Thus, the compactness of the boiler design, rather than the heat transfer coefficient, is the major consideration in determining the pitch/diameter ratio. However, there is a narrowest gap between tubes, depending on the particle size, below which a sharp drop in the heat transfer coefficient will occur. Determination of this narrowest gap requires further experimental studies.

Secondary Variables

Fines Recycle. Elutriated fines can be recycled back to the original

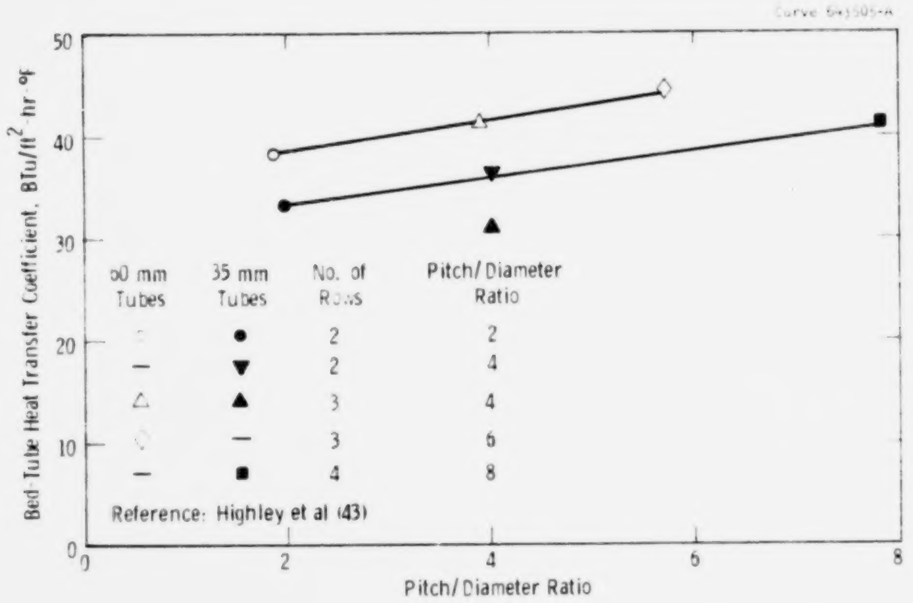

Fig. 1-49 – Change of bed-tube heat transfer coefficient with pitch/diameter ratio

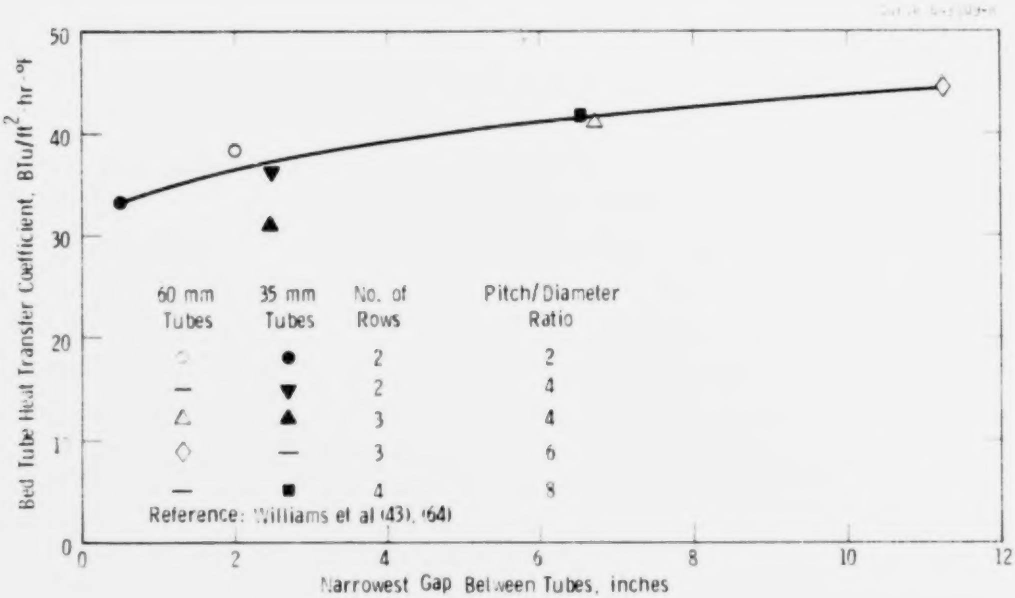

Fig. 1-50 – Change of heat transfer coefficient with tube spacing

bed or to the carbon burn-up cell. The fines recycle affects the heat transfer coefficient in two ways: recycle of fines changes the particle size distribution in the bed; large recycle of fines at a temperature different from that of the bed may create a local temperature gradient in the bed.

Coal Characteristics. Coal of different swelling numbers will change the particle size distribution in the bed after combustion. Different ash characteristics (elutriated or retained in the bed) will also change the particle size distribution. In general, the effect should be small, because coal represents <0.5% of the total bed material. For coal whose ash will accumulate in the bed after combustion, the effect will be more important. Caking properties of the coal will probably be important, too.

Air Distributor Design. Different air distributor designs will generate air bubbles of different sizes and frequencies and will cause different pressure drops. This affects the quality of fluidization, which is directly related to the effectiveness of mixing and heat transfer in the bed.

Unit Size. Slugging and poor mixing in the bed and, thus, poorer heat transfer, are more likely to occur in smaller units; larger units are relatively free of these problems. However, larger units with improperly designed air distributors are apt to cause by-passing of fluidizing air and affect the quality of fluidization. The major problem associated with this difference in unit sizes arises from applying the data on heat transfer coefficients obtained in smaller units to the design of larger units.

Bed Material. Bed material of different heat capacities will affect the heat transfer coefficient ($h \propto C_s^{0.25}$ to $C_s^{0.8}$). PER found the same heat transfer coefficients for a limestone bed and an ash bed, i.e., 47 Btu/ft^2-hr-°F [75].

Summary

Data on heat transfer coefficients obtained in fluidized bed combustion experiments by OAP contractors and NCB laboratories were presented in Table 1-5. The effect of changing different primary variables on heat transfer coefficients in the bed and the possible mechanisms of these effects are summarized in the following matrix. From the information shown in Table 1-5 and from the estimation (based on this matrix) of possible changes in heat transfer coefficient due to changes in operating and design variables, a heat transfer coefficient of 50 Btu/ft^2/hr/°F was chosen for design of the industrial, the atmospheric pressure, and the pressurized fluidized bed boilers. This heat transfer coefficient is considered reasonable and may be somewhat conservative in light of presently available information.

Heat transfer in fluidized bed combustors can be understood qualitatively with the data presently available, but quantitative design correlations are still lacking. Although considerable effort has been spent on understanding the heat transfer properties in the atmospheric pressure fluidized bed combustors, no comprehensive study has been performed for pressurized combustors with deep beds to investigate the effect of particle segregation and possible temperature gradients in the bed.

More data must be collected to confirm the meager information presently available on the effects of fluidizing velocity and tube arrangement in the beds, especially at higher fluidizing velocities (up to 15 ft/sec), smaller tube pitch/diameter ratios, and larger numbers of rows of tubes.

Scaling-up is still the most important problem for designing full-scale fluidized beds from the data accumulated in laboratory experimentation. Until the mechanics of gas-particle motion in a fluidized bed and its relation to the operating and design variables are better understood, no theoretical design correlation describing the heat transfer properties of fluidized beds can be formulated. The solution to

EFFECT OF DIFFERENT PRIMARY VARIABLES ON HEAT TRANSFER COEFFICIENT IN THE BED AND THE MECHANISMS INVOLVED

VARIABLE	CHANGE	HEAT TRANSFER COEFFICIENT	MECHANISM
Bed temperature	Increase	Increase	Increase in thermal conductivity and thermal diffusivity; increase in contribution of radiation
Particle size	Decrease	Increase	Increase in particle mobility and decrease in thickness of the gas film
Fluidizing velocity	Increase beyond the minimum fluidizing velocity	First increase rapidly then decrease slowly	Increase in particle agitation Decrease in particle concentration in the bed
Geometric factors			
walls or immersed tubes	--	Higher to the immersed tubes	Bubbling phenomena and solid circulation pattern in the bed
vertical or horizontal tubes	--	contradictory results	Possible particle deposit on upper half of the horizontal tubes; possible channeling effect for vertical tubes; tube pitch/diameter ratio may be important
tube bank in different bed depth	--	No change. Higher close to surface of the bed	At good fluidization. Segregation of particles in the bed results in sluggishly fluidized particles at bottom.
in splashing zone	Cold experiment hot experiment	30% drop from that in the bed. No change	Decrease in particle concentration. Radiation contribution; additional combustion above the bed.
tube pitch/diam.	Decrease	Decrease	Decrease in particle mobility
number of staggered tube rows	Decrease	Decrease	Decrease in particle mobility
square or triangular pitch	--	Not clear	

the problem of scaling-up will mostly rely on the empirical or semi-empirical correlations obtained through small-scale laboratory experimentation.

Particle Carry-Over

For design purposes, knowledge of the rate of entrainment of solids from a fluidized bed, of the size distribution of the elutriated solids in relation to that in the bed, and of the variation of both these quantities with the height of the freeboard is very important. In the case of fluidized bed coal combustion for air pollution abatement, where limestone or dolomite is recommended as bed material, the increase of dust loading from limestone fines and by limestone decrepitation and attrition may be expected to be the primary consideration for particulate control. When limestone is the bed material, the solids elutriated from the fluidized bed consist of coal ash, $CaSO_4$, unreacted CaO, and uncalcined $CaCO_3$. When dolomite is used, the elutriated solids will also contain the low reactive portion of MgO and $MgCO_3$.

Empirical investigations of the elutriation rate from an inert bed have been reviewed by Kunii and Levenspiel [60]. Studies of the elutriation rate from an ash bed were also conducted by Williams [78] in a one-foot square cold bed at 2.5 ft/sec fluidizing velocity. All studies show that the elutriation of the selected size fraction is directly proportional to its concentration in the bed and the excess fluidizing velocity (fluidizing velocity minus minimum fluidizing velocity - see Fig. 1-51), with the proportional constants being different for different materials and different particle sizes. These correlations are available [60, 78]. Thus it is possible to estimate the elutriation rate of a particular fresh limestone or dolomite once its attrition and decrepitation rates in a particular environment are determined. For limestones and dolomites which repeatedly go through absorption-regeneration cycles, the attrition and decrepitation rates are much more difficult to determine and will depend very much on their temperature history and on the reactor operating conditions. For the elutriation of carbon fines, the situation is also complex because the particles are no longer inert; the elutriation of carbon fines depends now not only on their concentration in the bed and the fluidizing velocity but also on bed temperature, excess air, and their relative combustion rate in the particular operating conditions.

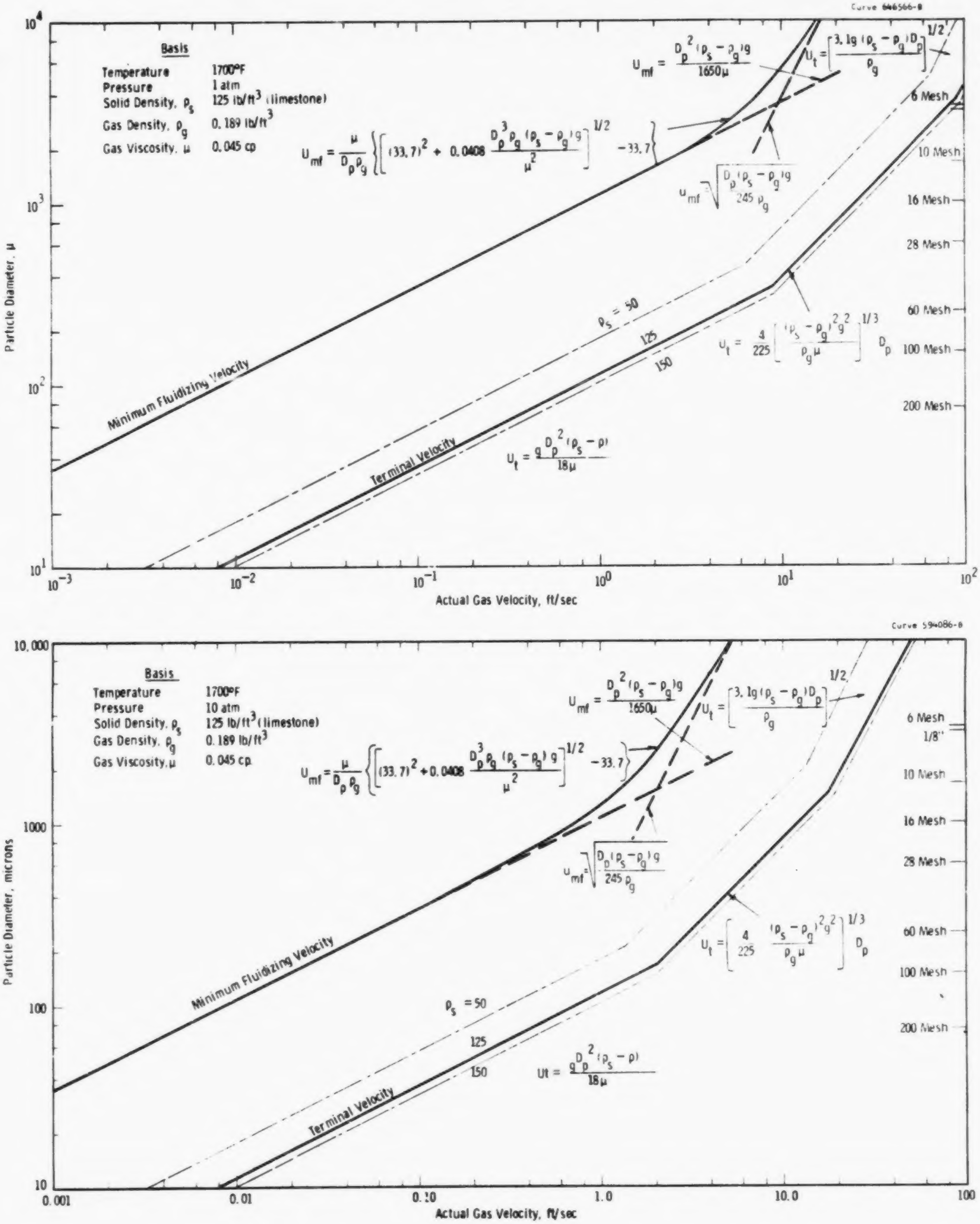

Fig. 1-51 – Fluidizing velocity curves for 1 atm. and 10 atm. operation

The experimental data for elutriation of carbon fines were correlated and discussed in the Combustion Efficiency section. Our primary concern here is to determine the rate of elutriation or dust loading at different operating conditions and their respective particle size distributions.

Since the evaluation of the combustion of coal in different beds (ash, refractory, or limestone bed) shows no clear indication that the type of bed material will affect the coal combustion efficiency, it is considered reasonable to assume that the elutriation of ash and limestone are independent processes and that their elutriation rates are additive. For certain coals with ash content less than ∿ 15%, notably the Ohio Pittsburgh No. 8 seam, experimental evidence [15] shows that most of the ash is elutriated out during operation. Thus the total elutriation rate is the rate of ash input from coal (which can be determined accurately through coal analysis) plus the independent elutriation rate of limestone or dolomite (which can be experimentally determined for each stone at specific operating conditions).

The elutriated limestone or dolomite comes from three sources: 1) attrition of the particles in the bed followed by subsequent elutriation; 2) decrepitation of particles due to their history-regeneration process, temperature cycles, etc.; 3) the size fraction of the feed material which has terminal velocities lower than the operating velocity. Unfortunately, very few data are available in these areas, except the limited attrition data from Esso, PER, and Consol[37, 75, 84, 107] obtained at low velocities and in small-batch laboratory units.

The picture is also confused by the fact that the physical integrity of the stone depends very much on the origin, preparation and temperature history of the individual stone. Argonne [3] reported in their studies that after calcination at 1600°F of once-screened 30 μ dolomite 1337 particles, 40% of the material decrepitated to fine powder. Esso [84] studied the rate of attrition, followed by elutriation, of calcined dolomite 1337 and limestone 1343 and 1359 up to excess superficial velocity of 5 ft/sec. They found that the attrition rate of dolomite was up to eight times larger than that of tested limestone. Assuming that extrapolation of Esso's data to higher velocities is possible, a fluidized bed of limestone 1359 particles of average size ∿ 2500 μ and running at 15 ft/sec will have an attrition rate of ∿1.5% of bed weight per hour; for dolomite 1337 this would be 12% of bed weight per hour. In another study [31], where fine coal (200 μ) was burned in a coarse limestone bed (930 μ), the stone 1359 elutriation rate was found to decrease from the initial particles "rounding off" period of ∿ 5% bed weight per hour to < 1%/hr in about two hours.

PER [75] studied the attrition of limestone 1359 in a batch reactor of 12 in x 16 in at fluidizing velocities 8 to 12 ft/sec and found the attrition rate during calcination was 5% to 7% of initial charge per hour; during sorption 3% per hour; and during regeneration 4% per hour. In Consol's studies [107] (superficial velocity 2 to 3 ft/sec), where Tymochtee dolomite was used, the attrition rate for highly sulfated dolomite was about 0.5% per hour. However, for lightly sulfated dolomite, an attrition rate as high as 4% per hour was noted. In addition, the size degradation was also found to depend on the type of coal or char burned in the fluidized bed. Thus, in a coarse bed operation, where the terminal velocities of all the particles are greater than the operating velocity, the rate of elutriation can be assumed to be the rate of attrition alone. To provide this narrower particle size distribution, the as-crushed material usually has to be single- or double-screened. The available elutriation data obtained by OAP contractors and British laboratories are summarized on page 130.

At steady-state, once-through operation where the bed material is continuously fed and withdrawn, the elutriation rate will depend not only on the bed weight but also on the material feed rate as well, if the feed material consists of fine particles. In addition to the attrition followed by elutriation, the size fractions of the feed material which have terminal velocities smaller than the operating fluidizing velocity will also be elutriated. The total amount of these fractions elutriated will depend on the residence time of the particles in the bed or the change-over rate of the bed material, because part of these solids will be withdrawn from the bed rather than elutriated from the top. The worst case here would be that in addition to the attrition and decrepitation contributions, the size fractions with terminal velocities smaller than the operating fluidizing velocity are completely elutriated.

When spent stones are to be regenerated and recycled with make-up stones, the attrition in the regeneration and in the pneumatic transport lines also has to be taken into account. The temperature cycling of the stone particles by going back and forth between the combustor and regenerator will also affect the physical integrity of the particles. Studies by Esso [31] on 1359 limestone at 2 to 3 ft/sec velocity showed little decrepitation up to six absorption - regeneration cycles. Previous studies [84] in a similar system using the same stone showed that less than 0.5% bed weight per cycle was elutriated after the first cycle (3.5% for the first cycle). How to extrapolate these data to higher velocities in a continuous absorption-regeneration operation is not clear; more studies should be done.

Changing the fluidizing velocity usually changes not only the rate of elutriation but also the size distribution of the elutriated solids. When fluidizing velocity is increased, the smaller-size fractions will change little because of their low terminal velocities. However, the larger-size fractions will progressively increase in amount with increasing velocity so that the weight fraction of the smaller particles tends to decrease and that of the larger particles tends to increase, as shown in the studies by BCURA in a 27 - in diameter rig [42] (Figure 1-52). In addition, the smaller-size fractions tend to contain more unburned carbon than the large-size fractions. As the fluidizing velocity increases, the carbon content in elutriated solids increases accordingly, because of decrease in residence time of the coal particles. But the carbon content in the large-size fractions tends to increase more rapidly than that in the smaller-size fractions (Figure 1-52). This may be because of two factors: first, the amount of the smaller-size fractions elutriated changes little with the fluidizing velocity; and second, smaller-size particles have higher combustion rates due to their larger surface area/weight ratio.

In a steady-state operation, the elutriation rate and the elutriated particle size distribution depend on the size distribution of feed material as well. To be meaningful, the size distribution of feed has to be specified. In the industrial application, the feed material can be fed as crushed or it can be single- and double-screened, depending on various application considerations. The end product of these operations may be a vastly different size distribution. The typical as-crushed coal or limestone size distribution from a Koppers reversible hammermill designed for crushing 5 in x 0 to 95% through 1/4 in with a minimum of minus 100 mesh is shown in Figure 1-53, along with the size distribution of coal feed used by different research organizations. With the coal feed size distribution as shown, the elutriated particles from the combustors at BCURA (27 in - diameter unit) and PER (18 in x 6 ft unit) using -1/4 in coal and at 10 to 15 ft/sec fluidizing velocity give comparable size distribution of elutriated solids (Figure 1-54). The elutriated particle size distribution from the carbon burn-up cell shown in Figure 1-54 is based primarily on PER's experiments [76], where the fly ash of 30% to 60% carbon content with a size distribution comparable to the solids elutriated out from the primary combustor was injected into the simulated carbon burn-up cell. It is cautioned that the data presented here are obtained from combustion of coal in either an ash bed or an inert bed such as refractory. When limestone or dolomite is used as bed material, the rate of elutriation and the particle size distribution of the elutriated particles will depend on the type of stone used -- limestone or dolomite of different classifications; operational factors -- whether the stone

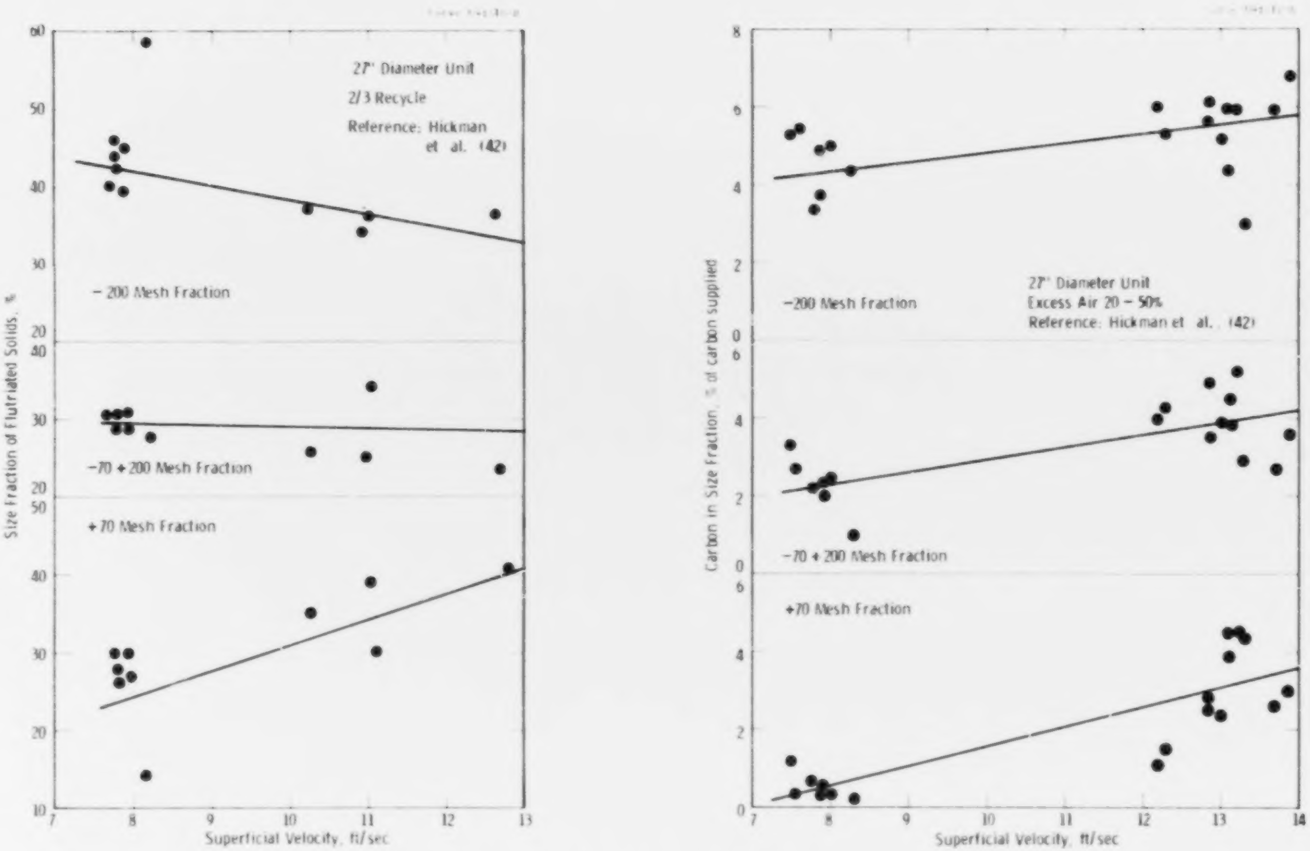

Fig. 1−52 − Effect of fluidizing velocity on particle size and carbon distribution of elutriated solids

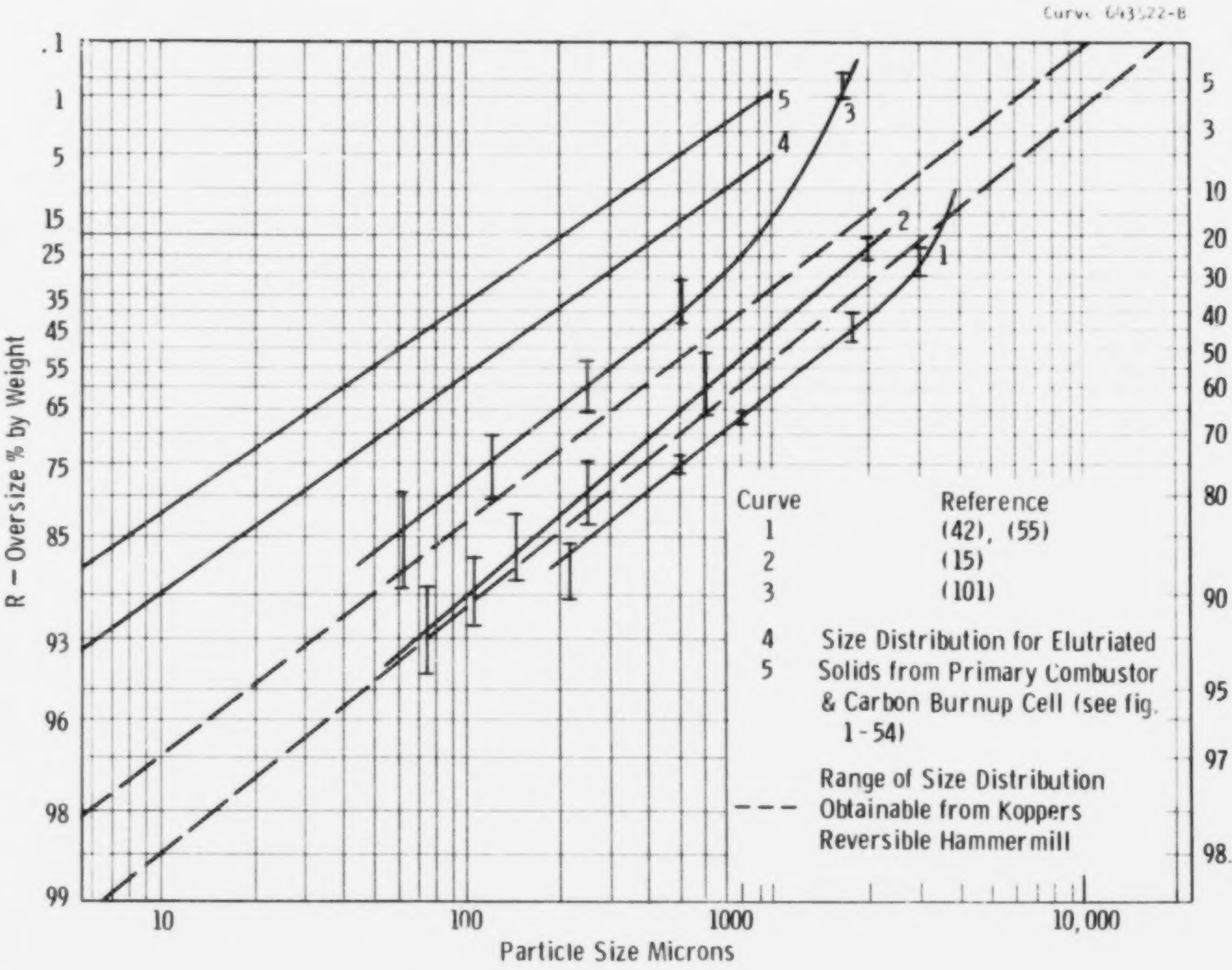

Fig. 1-53–Size distribution of coal feed

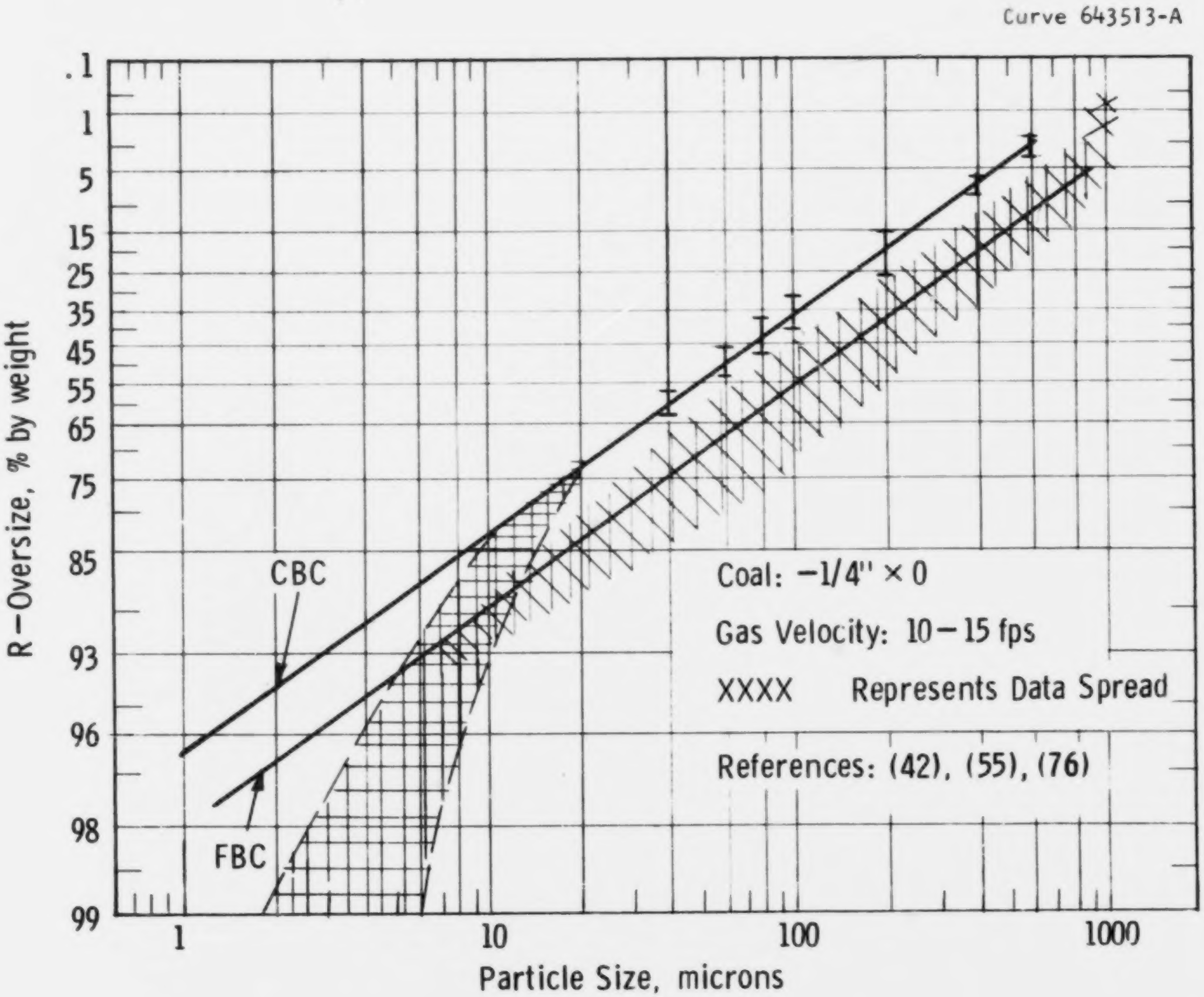

Fig. 1-54–Projected particle size distribution for material elutriated from fluidized bed combustor (FBC) & carbon burn-up cell (CBC)

will be used once-through, regenerated on-site and recycled, or regenerated off-site; and particle size distribution of the feed stones--as-crushed, single-, or double-screened.

For example, if as-crushed stone with a size distribution similar to that shown in Figure 1-53 (from Koppers' reversible hammermill) is used once-through in a fluidized bed operating at 15 ft/sec, the size fractions with lower terminal velocities can be elutriated out. The total amount which will be elutriated depends on the changeover rate of the bed material, with the maximum rate ∿ 35% of the total feed rate at the size distribution and fluidizing velocity specified. In the absence of experimental data, the particulate removal equipment can be designed for the most pessimistic case, where all-size fractions with terminal velocities lower than the operating velocity are completely elutriated - that is, 35% of the total feed rate in the present example. When the as-crushed stones are merely used as make-up supply for the regenerated stones, the elutriation rate can be taken to be 35% of the make-up stones plus the attrition rate of the recycled stones. For example, if the make-up stones constitute 10% of the total stone input, and the attrition of the recycled stones represent 1.5% of the total weight, then the total elutriation rate would be about 5% of the total stone feed rate. Hence, the use of as-crushed stones for once-through application will generally give much higher dust loading than the regeneration case when the attrition rate of the stones is small. In this case, single- or double-screening the as-crushed stones will effectively cut down the dust loading. When the specified particle size distribution is such that the terminal velocities are larger than the operating velocity, the elutriation represents the most optimistic case and is only due to attrition of particles in the bed.

No elutriation data are available for combustion of coal in limestone or dolomite beds operating at high velocities, in larger units, or in continuous absorption-regeneration operations. Further work is needed to provide the required data for accurate estimation of the elutriation rate for design purposes.

Summary

When limestone or dolomite is used as bed material, the increase in dust loading from stone fines and by stone attrition and decrepitation may be expected to be the primary consideration for particulate control. The design of particulate removal equipment is also complicated by the fact that the physical integrity of the stones depends very much on the origin, preparation, and temperature history of the individual stone. Unfortunately, most of the elutriation studies performed so far have been confined to batch combustion of coal in ash beds at low velocities. The available information is summarized in the following matrix. Based on the available information, design bases were chosen for preliminary design of the industrial, atmospheric pressure, and pressurized fluidized bed boilers as shown in the second matrix. Data are needed from combustion of coal in limestone or dolomite beds operating at high velocities (up to 15 ft/sec), in larger units, and in continuous absorption-regeneration operations.

SUMMARY OF AVAILABLE ELUTRIATION DATA AT ATMOSPHERIC
PRESSURE FROM [illegible] CONTRACTORS AND [illegible] LABORATORIES

ORGANIZATION	REACTOR SIZE	BED MATERIAL	PARTICLE SIZE	BED TEMPERATURE (°F)	FLUIDIZING VELOCITY (FT/SEC)	ELUTRIATION OR ATTRITION RATE	REFERENCE
USBM	18" diameter	refractory	Bed:-16+48 mesh coal:1/8" x 0	1450-1650	2.4-3.4	Refractory: 0.1% of bed weight per hour. ash: 50%-100% for 6 different coals.	[15]
Argonne	6" diameter	alumina	Bed: 30 mesh coal: -14 mesh sorbent: 25μ 100μ, 300μ	1600	3	Ash: 100%; 80%-90% of sum of ash and sorbent with 100μ sorbent; 60-80% with 300μ sorbent.	[3]
Esso	3" diameter	1337 1343 1359	Coarse pasticks for studying rate of attrition	Room temp.	Excess superficial velocity: 0-5	Attrition Rate: 0-3.5% bed weight/hr for BCR 1337 0-0.5% bed weight/hr for BCR 1343 0-0.7% bed weight/hr for BCR 1359	[84]
Esso	3" diameter	1359	Bed: 930μ coal: 200μ	Combustion: 1600 regeneration: 2000	Combustion: 3 Regeneration: 2	Rounding-off period: 5% bed weight/hr steady state: <1% bed weight/hr cyclic absorption-regeneration: 3.5% for 1st cycle; <0.5% for 2nd to 10th cycle.	[31]
PER	12" x 16"	1359	Bed:-10+20 mesh coal: 1/4" x 0	Combustion: 1500-1600 regeneration: 1900-2000	8-12	Calcination: 5-7% initial charge/hr sorption: 3% bed weight/hr regeneration: 4% bed weight/hr	[75]
Consol$^{(a)}$	4" diameter	Tymochtee Dolomite	-16+28 mesh	1700-1900	2-3	0.5% bed weight/hr for highly sulfated dolomite; 4%/hr for lightly sulfated dolomite.	[107]
NCB-CRE	12" square	Ash	-10 mesh	Room temp.	2.5	R = KC R = elutriation rate C = concentration of fines in the bed K = a proportional constant depends on operating velocity and terminal velocity of fines fraction.	[78]

NOTE: USBM = U.S. Bureau of Mines at Morgantown, West Virginia
Argonne = Argonne National Laboratory
Esso = Esso Research and Engineering Company, Linden, New Jersey
PER = Pope, Evans, and Robbins, Consulting Engineers, Alexandria, Virginia
Consol = Consolidation Coal Company
NCB-CRE = National Coal Board - Coal Research Establishment, England

(a) Operating pressure 8 psig.

DESIGN BASES FOR PARTICULATE REMOVAL EQUIPMENT FOR THE INDUSTRIAL, ATMOSPHERIC-PRESSURE, AND PRESSURIZED FLUIDIZED BED BOILERS

	INDUSTRIAL BOILDER	ATMOSPHERIC-PRESSURE UTILITY BOILER	PRESSURIZED UTILITY BOILER
ASH	100% elutriated	100% elutriated	100% elutriated
SORBENTS (-1/4" with minimum of fines)	3% of limestone feed rate(a)	3% of limestone feed rate(a)	5% of limestone feed rate(a)
PARTICLE SIZE DISTRIBUTION			
From Primary Bed	Figure 1-54	Figure 1-54	Figure 1-54
From CBC	Figure 1-54	Figure 1-54	Figure 1-54

(a) Limestone feed rate is the combined feed rate of the recycled stream and the make-up stream. The recycled stream may contain a smaller amount of fines than that in the make-up stream but usually has higher attrition and decrepitation rates.

Boiler Tube Corrosion, Erosion and Fouling

Introduction

The economics and performance of fluidized bed combustion boilers with immersed heat transfer surface depend on the resistance to corrosion, erosion, and fouling of the heat transfer surface in the fluidized bed and in the freeboard. Theoretical analyses have been made to assess potential corrosion, erosion, and fouling of boiler tubes at projected operating conditions. Experimental data have been obtained on tubes in fluidized beds of limestone and dolomite burning various coals. An evaluation of the available theoretical and experimental information indicates that conventional boiler tube materials can be used in the proposed fluidized bed combustion boiler designs without encountering corrosion, erosion, or fouling problems.

The fluidized bed boiler may also enable power plants to operate at steam conditions in excess of present practice. In the case of conventional boilers, the steam cycles have hitherto been limited to a temperature of 1050°F because of fouling and corrosion of superheater tubes. Fluidized bed combustion of coal conducted at a much lower temperature (1400°F to 1800°F) compared to a flame temperature of up to 3000°F in conventional boilers promises to cut down fouling, corrosion, and erosion of the boiler tubes, and permits operation at higher-temperature steam cycles in addition to providing a higher heat transfer coefficient in the bed.

Theoretical Evaluation

Metal wastage in boiler tubes may occur in many different ways. It can occur from internal erosion and corrosion by steam and its impurities, such as Cl^- and O_2. This is especially true when low-alloy tubes are used. It can also occur through external abrasion, fouling, and corrosion. The internal water/steam-side corrosion/erosion has been pretty well characterized through the accumulation of experience in operating steam power plants.

Thus, the present discussion will be restricted to the fouling, corrosion, and erosion of tubes immersed in a fluidized bed. The operating conditions in the beds which affect corrosion, erosion, and fouling are summarized in Tables 1-6 and 1-7.

Erosion of metal caused by dust particles is generally a function of particle size, particle shape, particle physical properties, impact velocity, impact angle, and the metal physical properties, as discussed in the next section, on gas turbine blade erosion. Although the problem of erosion in the conventional coal-fired boiler is not as severe as that in the gas turbine, the ash erosion problem does exist in the conventional boilers and must not be ignored even though coal ash particles may be exceedingly fine. The "hard" nature of p.f. coal ash, cenospheres, makes the particles particularly erosive. Where ash particles are concentrated, as at turns formed by baffles within boiler banks, erosion is a potential problem which can be remedied by limiting gas velocities. The maximum allowable gas velocity recommended for the conventional pulverized coal-fired boiler is ∿100 ft/sec, and that for the coal-fired spreader stoker is ∿60 ft/sec [90].

External erosion of the immersed tubes in the fluidized bed may result from the vigorous particle motion which continuously scours the tubes' surface. Erosion per se is not considered important in the fluid bed application where the maximum gas velocity in the boilers is 15 ft/sec, and the particle velocity is considerably less than the gas velocity due to the high frequency of inelastic particle collisions in the bed. In addition, the vigorous particle motion may also be effective in scrubbing off deposits formed on the heat transfer surfaces. If the fouling rate is higher than the eroding rate, the particle erosion actually has an advantageous effect because it keeps scale from building up at a high rate.

Corrosion will occur predominantly by chemical attack with different mechanisms such as surface oxidation, removal of the protective scale on metal surfaces through chemical reactions with corrodents, and

TABLE 1-6

OPERATING CONDITIONS IN FLUIDIZED BED BOILER DESIGNS

	INDUSTRIAL BOILER	ATMOSPHERIC PRESSURE UTILITY BOILER	PRESSURIZED UTILITY BOILER
PRESSURE, ATM	1	1	10
BED TEMPERATURE, °F	1300-2000	1300-2000	1300-2000
SUPERFICIAL GAS VELOCITY, FT/SEC	≤ 13	≤ 10	≤ 10
MAXIMUM GAS VELOCITY BETWEEN THE TUBES IN THE BED, FT/SEC	15	24	14
EXCESS AIR IN PRIMARY BED, %	10	10	8.5
DESIGN MEAN TUBE WALL TEMPERATURE, °F			
Water walls	570	975	975
Pre-evaporator	-	710	732
Evaporator	570	850-890	975
Superheater	790	875-1175	900-1150
Reheater	-	1100	1122

Table 1-7

ANALYSIS OF COAL ASH, COAL, AND BED MATERIAL

COAL ASH (Same for all Designs)

Compound	Wt %
SiO_2	45.3
Fe_2O_3	27.3
P_2O5	0.11
MgO	0.60
K_2O	1.80
Al_2O_3	21.2
TiO_2	1.0
CaO	1.9
Na_2O	0.2
SO_3	0.7

COAL

Compound	Wt %
Chlorine	0.05
Sulfur	4.3

BED MATERIAL

Compound	Wt %	
	Limestone 1359	Dolomite 1337
SiO_2	0.85	0.78
Al_2O_3	0.30	0.15
Fe_2O_3	0.17	0.25
MgO	1.07	45.0
CaO	97.	53.0
TiO_2	<0.05	0.02
SrO	0.07	<0.03
Na_2O	<0.02	<0.02
K_2O	<0.1	<0.1
MnO_2	<0.05	<0.03

direct chemical attack on the metal surfaces. The corrosion of the immersed tubes in the fluidized bed is expected to be much less severe than that obtained on the steam tubes of modern, high-temperature p.f. boilers, due to the relatively low bed temperature employed. Figure 1-55 shows the variation with temperature of the vapor pressure of some of the alkali metal salts usually present in coal ashes [9,28,45,79]. The vapor pressure of the salts at normal flame temperatures of ∿2500°F is several orders of magnitude higher than that in a fluidized bed operating at 1400°F to 1800°F. Also, the only alkali compounds having significant vapor pressure at the fluid bed operating temperature are the chlorides and oxides. Therefore, it is considered unlikely that carbonates or sulfates, which are common causes of fouling and corrosion in conventional boilers, will be a problem. In addition, an operating temperature of 1400°F to 1800°F is considered too low to allow formation of sintered ash deposits on the boilers through reactions of the volatilized alkali compounds with aluminosilicates. However, deposits of alkali chlorides on heat transfer surfaces in fluid bed combustors burning high chlorine content coals can lead to corrosion. Chlorine has been known to contribute significantly to the fouling and corrosion tendencies of coal when present in concentrations higher than 0.3% [79].

The theoretical analyses indicate that corrosion, erosion, and fouling will not be a problem for the proposed design conditions and that higher steam temperatures are practical for fluidized bed boilers.

Experimental Data

Experimental data on corrosion, erosion, and fouling of boiler tube materials are available from work conducted in England [34,69,70,71] and the United States [76,97]. Work sponsored by OAP is being continued by NCB and the U. S. Bureau of Mines, Morgantown, West Virginia. A summary of the experimental work is presented in Table 1-8.

Dainton and Elliott [22] immersed a stainless steel probe in a fluidized bed burning Thoresby coal of high chlorine and sulfur content

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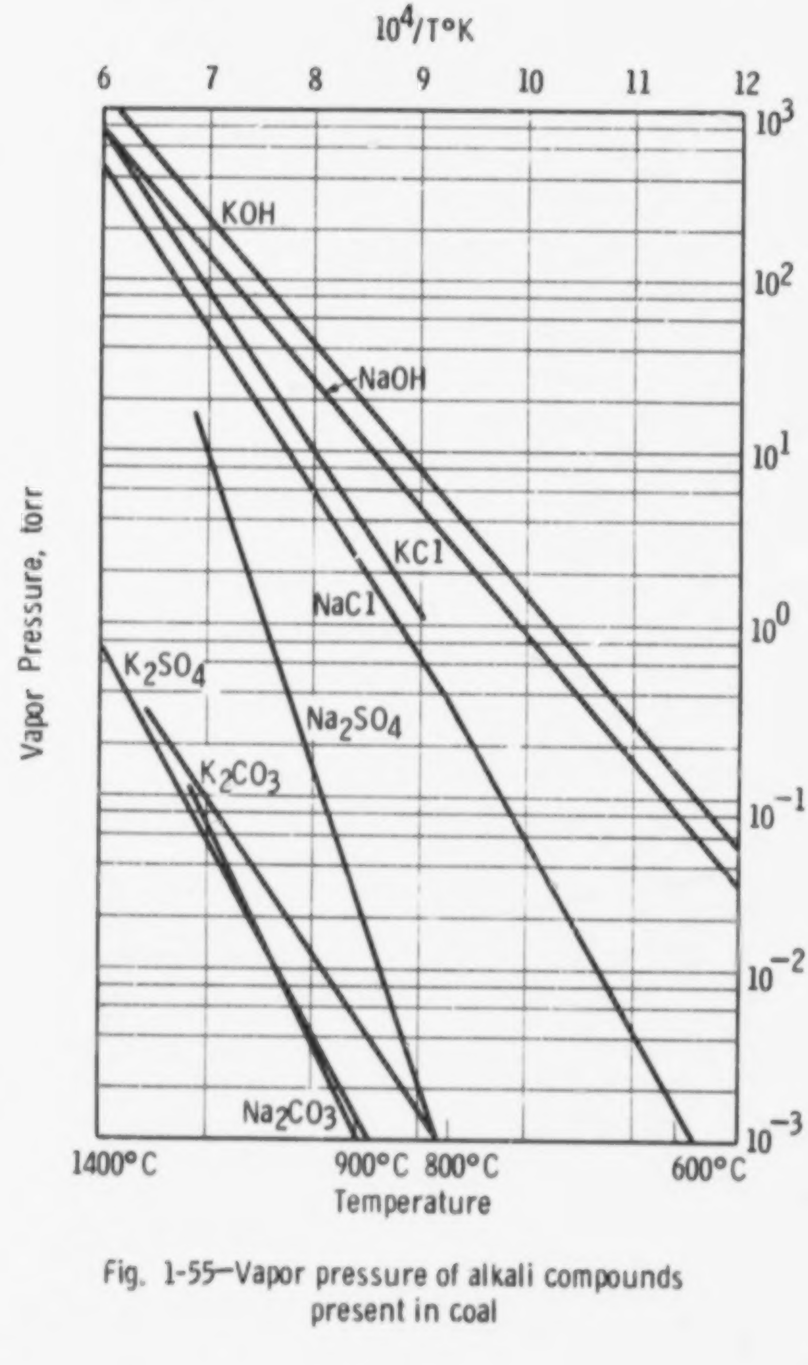

Fig. 1-55—Vapor pressure of alkali compounds present in coal

TABLE 1-8

EXPERIMENTAL WORK ON FLUIDIZED BED BOILER TUBE MATERIALS

INVESTIGATOR	APPARATUS	OPERATING PRESSURE & FLUIDIZING VELOCITY	BED TEMP. °F	METAL TEMP. °F	NUMBER OF MATERIALS STUDIED	NUMBER OF COALS STUDIED	NUMBER OF LIMESTONES STUDIED	DURATION OF TESTS, HRS.
Dainton & Elliott [22]	8" dia.	1 atm	1300	900-1200	1	1	0	120
Gliddon [34]	8" dia.	1 atm	1300	930-1110	1	8	0	30-150
NCB [69, 70, 71]	48" x 24"	3-5 atm 2 ft/sec	1470	620-1470	7	2	2	100
	12" dia.	1 atm 3 ft/sec	1560	735-1290	6	1	1	100-500
	27" dia. (test not completed yet)	1 atm 8 ft/sec	1522	750-1290	6	1	1	500-1000
PER [76]	72" x 18"	1 atm 12-15 ft/sec	1600	~600	(not primarily intended for corrosion studies)			
USBM [97]	18" dia.	1 atm	1500-1600	600-1200	4	(work not started yet)		

(Cl = 0.86 wt%; S = 1.06 wt%) at a temperature of 1300°F for 120 hours and found that the metal surface at a temperature above 1100°F was free of any deposit or corrosion. The fouling became progressively worse when the metal surface temperature decreased below this level, due to the increased temperature differential between the probe and the bed. The deposit was predominately sodium chloride. The same coal burned in a test rig at BCURA simulating p.f. firing gave massive deposits on superheater tubes which almost bridged across the tube bundle in a 24-hour experiment [28]. In the same series of experiments, Gliddon, of the Central Electric Generation Board, England (CEGB) [34], combusted four coals and four chars -- Thoresby, Babbington, Pye Hill, and Ollerton -- of different sulfur and chlorine contents. He found that the maximum fouling deposition rate for the high-chlorine Thoresby coal (0.86 wt% Cl, 1.06wt% S) was approximately six times that for the Babbington char (0.33 wt% Cl, 0.48 wt%S). The majority of the deposit for Thoresby coal is sodium chloride, and the rate of deposition falls off sharply at probe metal temperature above ∿950°F, suggesting that the dew point of sodium chloride at the operating condition is between 950°F and 1100°F. This may necessitate special start-up and shutdown procedures to prevent deposition of sodium chloride. Traces of sulfate were found in Babbington deposits but none in any of the Thoresby deposits. The results also show that the rate of fouling can be qualitatively predicted by the vapor pressure curve of sodium chloride. The typical fouling rate for Thoresby coal is presented in Figure 1-56. A 30-hour test with Pye Hill coal (0.42 wt% Cl, 2.71 wt% S) indicated no visible corrosion of the probe. The probe also showed no marked signs of corrosion after 150 hours in the bed burning Ollerton coal (0.3 wt% Cl, 1.33 wt% S), although a very thin adherent layer of scale observable under magnification was forming in places along the probe. The deposits contained significants of alkali metals, but no chloride or sulfate was detected. The experimental data indicate that the initial rate of attack is often much higher (about 8.5 x 10^{-3} inches per 1000 hours for Thoresby coal) than that which occurs after a long period of operation under constant conditions.

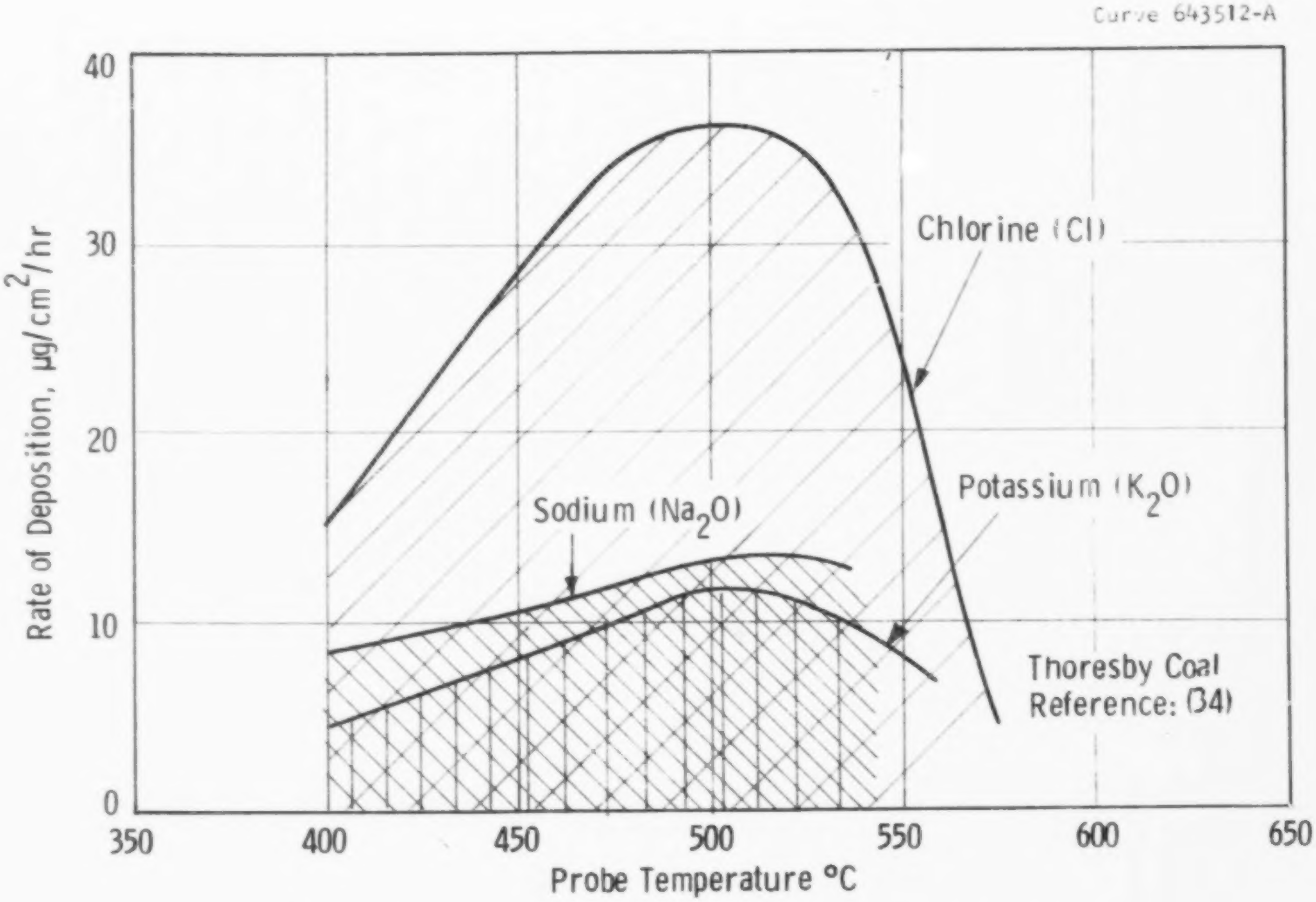

Fig. 1-56—Effect of probe temperature on rate of deposition

The most comprehensive and systematic studies on corrosion and deposition of boiler tube materials are being conducted by NCB under a contract with OAP [69,70,71]. The corrosion/erosion studies are divided into three parts: high-pressure operation, short-duration atmospheric pressure operation at low velocities, and long-duration atmospheric pressure operation, to assess the effect of limestone addition.

Corrosion/Erosion of Boiler Tube Materials under Pressurized Operation Conditions. The investigation was performed by combusting a U. K. coal with a U. K. dolomite and by combusting a U. S. coal with a U. S. dolomite in a 48 in x 24 in pressurized fluid bed at 3 to 5 atm pressure. The U. K. coal chosen was Welbeck coal, with 0.53 wt% Cl and 1.25 wt% S; and the U. S. coal was Humphrey No. 7, with 0.08 wt% Cl and 2.75 wt% S. Specimens of seven boiler tube materials (Table 1-9) formed into continuous tubes were immersed in the bed and air-cooled to the temperature range 620°F to 1470°F. The test specimens were descaled and weight losses examined after each run of ∿100 hours duration at a fluidizing velocity of two ft/sec. The average weight loss for the U. K. coal and dolomite experiments ranged from 2 to 445 $\mu g/cm^2$ hr, depending on the temperature of the specimen and the specimen tube material. The weight losses for the U. S. coal and dolomite test were 9 to 361 $\mu g/cm^2$ hr. The detailed data are presented in Table 1-10. In all cases, visual observation indicated that the color, adherence, and type of deposits depended on the material and temperature of the specimens. Metallographic examinations found no evidence of intergrannular penetration in any of the tested specimens, and the higher weight losses of the 2-1/4% Cr-1% Mo, 1% Cr-1/2% Mo ferritic specimens were associated with a general surface roughening. The nonuniformity of the weight losses can, in some instances, be related to the position of the specimens relative to the coal feeding nozzles.

Corrosion Studies of Boiler Tube Materials at Atmospheric Pressure and Low Fluidizing Velocities. Three series of test runs were planned, two of which have been completed. U. S. Humphrey No. 7 coal was combusted in a 12-in rig at three ft/sec fluidizing velocity with

TABLE 1-9

TYPICAL ANALYSES OF METAL SPECIMENS

Designation	NOMINAL COMPOSITION %							
	Cr	Ni	Mo	Mn	Ti	Al	Nb	Fe
12% Cr	12							88
Rf 36	18	12					1	69
SF 316	17	12	2.5	2				66.5
Esshete 1250	15	10	1	6			1	67
P.E. 16	18	37	5		1.2	1.2		37
1%Cr 1/2%Mo	1		0.5	0.5				98
2 1/4% Cr 1%Mo	2.25		1	0.5				96.25

TABLE 1-10

AVERAGE TUBE SPECIMEN WEIGHT LOSS UNDER PRESSURIZED OPERATION CONDITION, mg/cm^2-hr.

TUBE MATERIAL TEMPERATURE RANGE, °F(a)		12% Cr		RF 36		SF 316		PE 16		Esshete 1250		1% Cr 1/2% Mo		2 1/4% Cr 1% Mo	
Series B(b)	Series U(c)	B	U	B	U	B	U	B	U	B	U	B	U	B	U
1350- 1460	1470 1450	-	-	16	145 63	15	166 80	7	50 37	50	223 207	-	-	-	-
1320- 1380	1280- 1380	5	18	8	48	8	54	5	36	28	210	-	-	-	-
1040- 1220	1100- 1180	5	15	4	28	4	23	3	13	5	48	-	-	-	-
730- 920	830- 1090	6	29	13	29	11	33	6	17	9	-	373	254	445	319
660- 990	620- 920	2	18	4	9	4	22	3	8	5	15	24	99	35	151

(a) Temperature ranges are the temperature differences between the ends of the composite tubes.

(b) Series B is the result obtained from combusting a U.K. coal and a U.K. dolomite.

(c) Series U is the result obtained from combusting a U.S. coal and a U.S. dolomite.

or without limestone addition. Test series 1 was run for 100 hours without limestone addition; test series 2, for 500 hours without limestone addition; test series 3, for 500 hours with limestone addition. Specimens of tube form and coupon were used in the bed and in the freeboard. The available results (test series 1 and 2) are summarized in Tables 1-11 and 1-12. There was no marked difference between these results and the similar tests under pressurized conditions reported above. The high chromium content alloys appeared satisfactory within the conditions tested. However, the medium carbon and 2-1/4% Cr-1% Mo steels suffered a higher rate of metal loss in 500 hours. This confirms the earlier findings described in the last section. In all instances, substantially less metal wastage was observed for the tubes and coupons located in the freeboard.

Comparing the present results with those from the earlier work burning a U. K. coal (Newstead coal with 0.6 wt% S and 0.3 wt% Cl) for 100 hours and 1000 hours, it is evident that most of the metal loss occurs in the first 100 hours. The variation of metal loss with operating time for these two coals is about the same and is shown in Figure 1-57. The comparison also implies that there is no measurable difference in the corrosive potential of these two coals.

Long-term Corrosion Tests at Atmospheric Pressure to Assess the Effect of Limestone Addition on Corrosion of Boiler Tube Materials. Four series of tests with duration of operation up to 1000 hours were run. At the same time, the particle size of limestone and the combustion conditions were changed to evaluate their respective effect on corrosion. The available data using U. S. Humphrey No. 7 coal without limestone addition at a fluidizing velocity of 8 ft/sec are presented in Tables 1-13 and 1-14. These tests showed that the medium carbon and 2-1/4% Cr-1% Mo ferritic steel specimens had suffered internal attack on the steam side, possibly by Cl^- and O_2 in the steam supply. Thus, a correction of internal weight loss is required. Again the data show that the austenitic steels perform satisfactorily; and of the chrome-ferritic steels, the 12% chromium steel behaves the best. The low-chromium steels suffer much higher weight losses.

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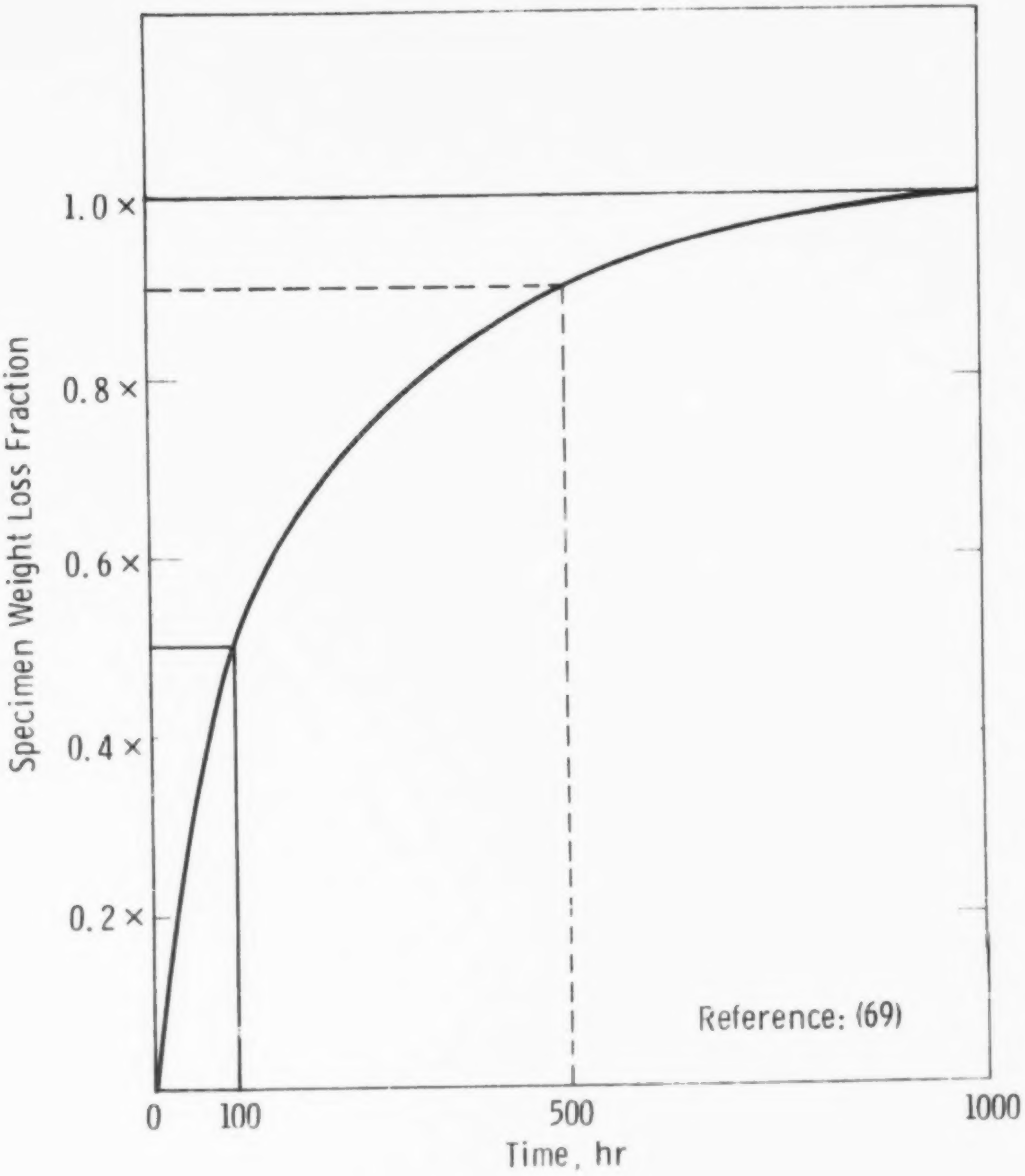

Fig. 1-57 —Effect of time on weight loss

(100 & 1000 hr rates: Newstead coal)
(100 & 500 hr rates: Humphrey No. 7 coal)

TABLE 1-11

SUMMARY OF RATE OF WEIGHT LOSS-μg/cm^{2}h
TEST SERIES 1--DURATION 100 hrs

LOCATION	NOMINAL TEMPERATURE OF SPECIMEN °F	RATE OF WEIGHT LOSS - μg/cm^{2}h					
		Type 347 Austenitic Steel	Type 316 Austenitic Steel	Esshete 1250	12% Cr Ferritic Steel	2-14% Cr Ferritic Steel	Medium C Steel
BED Tubes	750		1	1		67	74
			1	1		83	100
				3		109	90
						111	101
	923			3		153	125
			2	3		154	192
			2	3		198	143
						210	190
	1105	8	6	10		696	
		7	10	11		462	
		7	9	9			
			8	7			
	1275	10	9	17			
		12	13	21			
		11	13	24			
			11	24			
			11	19			
Coupons	1560	12	9	20			
FREEBOARD Tubes	730		1	2		86	63
				1		68	66
				2		65	62
						78	57
	1285	3	3	7			
		5	3	8			
		6	3	10			
			3	10			
			5	16			
Coupons	1500	3	2	4			

TABLE 1-12

SUMMARY OF RATE OF METAL LOSS-$\mu g/cm^2h$
TEST SERIES 2 -- DURATION 500 hrs

LOCATION	NOMINAL TEMPERATURE OF SPECIMEN °F	RATE OF WEIGHT LOSS - $\mu g/cm^2h$						
		Type 347 Austenitic Steel	Type 316 Austenitic Steel	Esshete 1250	12% Cr Ferritic Steel	2-1/4% Cr Ferritic Steel	Medium C Steel	Nimonic PE 16
BED Tubes	735		2	2	1	34	73	
			<1	<1	2	29	35	
				<1		33	36	
	930		2	2	5	81	65	
			1	2	8	64	67	
				1		87	132	
	1095	4	4	5		459		
		2	3	5	5	793		
			3	3		365		
	1290	3	5	6	2			3
		3	4	9	3			3
			4	2	5			5
Coupons	-	-	-	-	-	-	-	-
FREEBOARD Tubes	755		<1	<1	<1	29	36	
			<1	<1	2	23	21	
				<1		23	18	
	1285	1	1	2	1			1
		1	1	5	1			1
		2	2	6	3			2
Coupons	-	-	-	-	-	-	-	-

TABLE 1-13

OBSERVED RATE OF WEIGHT LOSSES ON SPECIMENS
DURATION 147 HOURS

NOMINAL TEMPERATURE (°F)	RATE OF WEIGHT LOSS ($\mu g/cm^2h$)					
	Type 347 austenitic steel	Type 316 austenitic steel	Esshete 1250	12% Cr ferritic	2 1/4% Cr ferritic	Medium C steel
750	8	14	7	21	248	A[(a)]
930	5	18	19	67	888	A
1110	11	13	20	38	A	A
1290	11	16	15	18	A	A

Note: These data may be affected to an unknown extent by the violent bed conditions, particularly temperature, pertaining to the shut-down.

(a) Specimens suffer internal attack on the steam side, possibly by Cl^- and O_2 in the steam supply, so data require additional correction.

TABLE 1-14

OBSERVED RATE OF WEIGHT LOSSES ON SPECIMENS
DURATION - 500 HOURS

NOMINAL TEMPERATURE (°F)	RATE OF WEIGHT LOSS ($\mu g/cm^2/h$)					
	Type 347 austenitic steel	Type 316 austenitic steel	Esshete 1250	12% Cr ferritic	2-1/4% Cr ferritic	Medium C steel
750	7	2	2	3	30	41
	4	4	2	4	53	55
	1		1	4	96	614
	8		5	4	111	80
			4			
			7			
930	7	4	6	6		301
	3	4	9	6	A[a]	A
	9	6	7	9		
			7	8		
			11			
1110	6	3	6	2		
		3	7	3		
		2	7	2		
		4	9	2	A	A
		4	8	5		
1290	13	6	11	3		
	3	8	12	4		
	8	8	12	4		
	4	14	21	4		
	12	13	22	5		
	15	10	19	6		

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[a] Specimens suffer internal attack on the steam side, possible by Cl^- and O_2 in the steam supply, so data require additional correction.

Although the complete data for this systematic study of corrosion/erosion on boiler tube materials are still not available, some conclusions can be drawn from the data summarized in Figures 1-58 to 1-63:

- The austenitic steels (Figures 1-61 to 1-63) give very small weight loss ($<10\ \mu g/cm^2$ after 500 hours operation) at all tested temperatures for lower chlorine content coals and at fluidizing velocities up to eight ft/sec. Increasing the fluidizing velocity increases the rate of weight loss.
- A high-chlorine coal is unlikely to cause corrosion of an austenitic steel heat transfer surface immersed in a fluidized bed if the metal temperature is kept above the dew point temperature of sodium chloride in the gas. Sodium chloride dew point temperature is determined mainly by the chlorine content of the coal and the bed temperature which determines its vapor pressure. The dew point would rise as the bed temperature is increased, thus increasing the temperature range over which metal corrosion is likely.
- The chrome-ferritic and medium carbon steels are more susceptible to attack, as shown by greater weight loss rates (Figures 1-58 to 1-60), although the rate tends to decrease with increasing operation time.
- Metallographic examination shows absence of intergranular penetration or pitting for tests now completed under normal operating conditions. Slight oxide formation is present in all instances.
- Rate of weight loss decreases with increase in operation time under constant operating conditions.
- Substantially less corrosion was found for the tubes or coupons situated in the freeboard as compared to that in the bed.

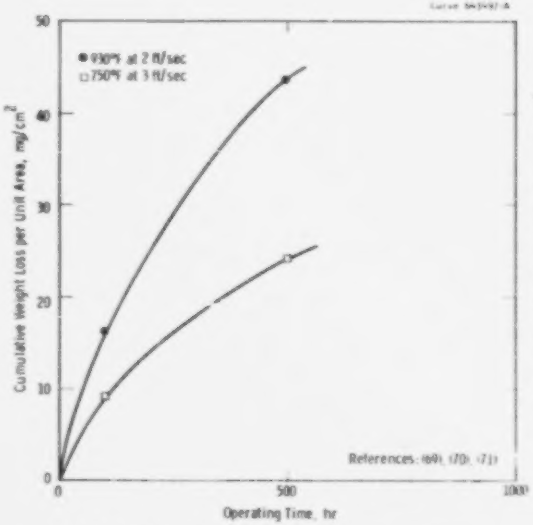


Fig. 1-58—Cumulative weight loss per unit area as a function of time for medium C steel

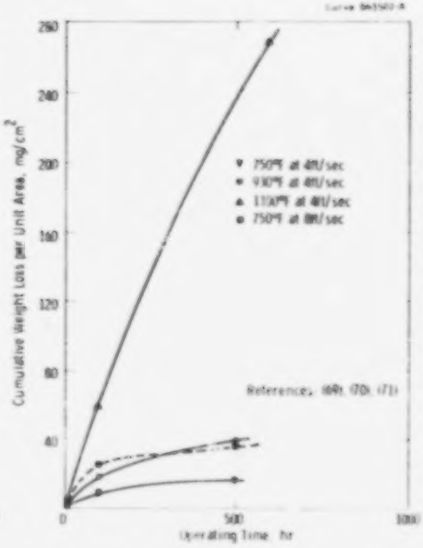


Fig. 1-59—Cumulative weight loss per unit area as a function of time for 2 1/4% chrome ferritic steel

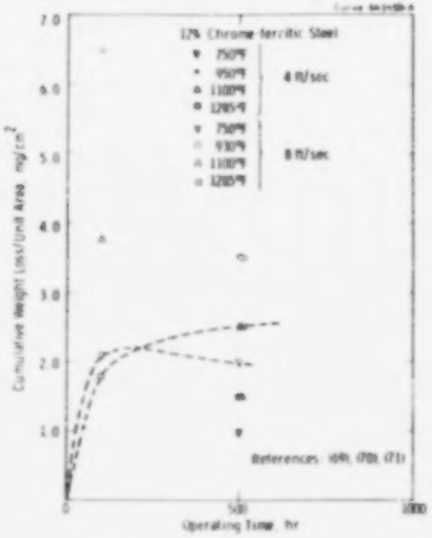


Fig. 1-60—Cumulative weight loss per unit area as a function of time for 12% chrome ferritic steel

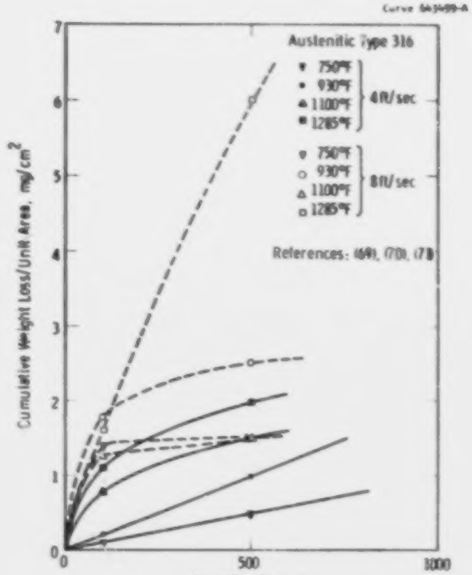


Fig. 1-61—Cumulative weight loss per unit area as a function of time for austenitic type 316 steel

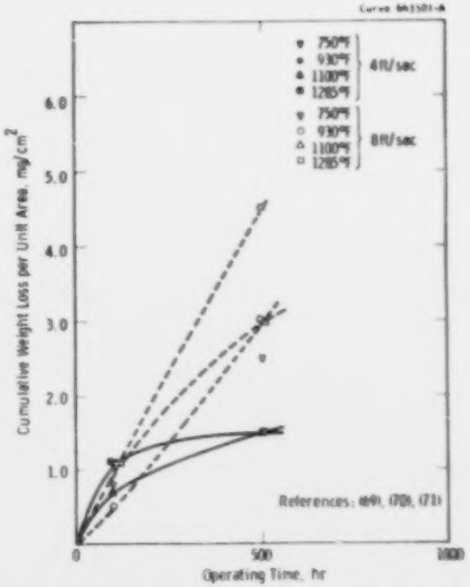


Fig. 1-62—Cumulative weight loss per unit area as a function of time for austenitic type 347 steel

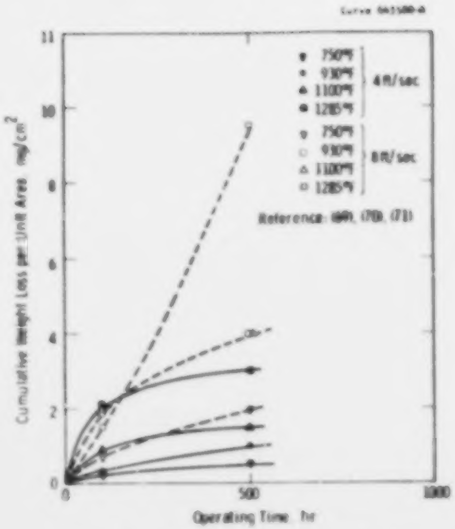


Fig. 1-63—Cumulative weight loss per unit area as a function of time for austenitic esshete 1250 steel

Boiler Tube Materials Selection

Boiler tube materials were selected for the present fluidized bed boiler designs on the basis of conventional practice. The selected tube materials and their respective chemical compositions are summarized in Tables 1-15 and 1-16. Since the materials were selected, results from the NCB have become available, and the materials selected will be reassessed when results from the NCB and USBM tests are complete. Preliminary evaluation indicates that the conventional boiler tube materials can be used in the fluidized bed application if the long-duration runs show that the rate of weight loss does indeed level off with increase in operation time. Final assessment can only be carried out after more data are accumulated at higher velocity, higher pressure, high temperature, and longer duration.

Assessment

Additional data on corrosion, erosion, and fouling of boiler tube materials are needed to evaluate critically the present fluidized bed designs.

- Longer-duration runs (up to 2000 hours) should be performed at a higher pressure (up to 10 atm), and a higher velocity (up to 15 ft/sec) to evaluate the rate of weight loss with respect to operation time.
- The effect of limestone addition and the effect of burning different coals with different impurities, especially the high-sulfur coals, should be carefully studied. The corrosion rate from burning a U. S. coal with a U. S. limestone has been up to seven times larger than the corrosion rate from burning a U. K. coal with a U. K. limestone (see Table 1-10).
- The effect of thermal cycling in the fluidized bed should be evaluated. The possibility of increasing corrosion due to deposition of sodium chloride during start-up and shut-down should be studied.

Table 1-15

FLUIDIZED BED COMBUSTION BOILER DESIGNS
BOILER TUBE MATERIALS

	INDUSTRIAL BOILER		ATMOSPHERIC PRESSURE BOILER		PRESSURIZED BOILER	
	Material	Design Temp. °F	Material	Design Temp. °F	Material	Design Temp. °F
Water walls	SA-178-A or SA-192	570	SA-213-T22	975	SA-213-T22	975
Pre-evaporator		-	SA-210-A1	710	SA-210-A1	732
Evaporator		570	SA-213-T2	850-890	-	-
Superheater	SA-213-T22	790	SA-213-T2	875-930	SA-213-T2	900
			SA-213-T22	1075	SA-213-T22	1058
			SA-213-TP-304H	1175	SA-213-TP-304H	1150
Reheater	-	-	SA-213-T22	1075	SA-213-T22	1122
			SA-213-TP-304H	1175		

TABLE 1-16

FLUIDIZED BED COMBUSTION BOILER DESIGNS
CHEMICAL COMPOSITION OF SELECTED BOILER TUBE MATERIALS

(in wt %)

	CARBON	MANGANESE	PHOSPHORUS	SULFUR	SILICON	CHROMIUM	MOLYBDENUM	NICKEL
SA-210-A1	0.27 max	0.93 max	0.048 max	0.058 max	0.10 min	-	-	-
SA-213-T2	0.10-0.20	0.30-0.61	0.045 max	0.045 max	0.10-0.30	0.50-0.81	0.44-0.65	-
SA-213-T22	0.15 max	0.30-0.60	0.030 max	0.030 max	0.50 max	1.90-2.60	0.87-1.13	-
SA-213-TP304H	0.04-0.10	2.00 max	0.04 max	0.03 max	0.75 max	18.00-20.00	-	8.00-11.00
welded tubing SA-178-A (low carbon steel)	0.06-0.18	0.27-0.63	0.05 max	0.06 max	-	-	-	-
seamless tubing SA-192	0.06-0.18	0.27-0.63	0.048 max	0.058 max	0.25 max	-	-	-

Gas Turbine Erosion and Corrosion

Introduction

A gas turbine is used in the fluidized bed boiler combined steam-gas turbine system to extract energy from the high-pressure, hot combustion gases leaving the boiler. The gas turbine in a typical application provides 15 to 20% of the net electrical output.

The gas which this turbine expands consists typically of 74 mole% N_2, 15 mole% CO_2, 2 mole% O_2 and 1.7×10^{-2} mole percent SO_2. It contains dust loadings of 0.1 to 0.15 grains per standard cubic foot. The dust composition is approximately 80% coal ash, 7% carbon, and 13% spent limestone sulfur sorbent. The anticipated particle size distribution is given in Table 1-17.

TABLE 1-17

ANTICIPATED PARTICLE SIZE DISTRIBUTION OF DUST ENTRAINED IN GAS TURBINE FEED

AVERAGE PARTICLE DIAMETER (MICROMETERS)	WEIGHT PERCENT IN SIZE RANGE WT %	CONCENTRATION GRAINS PER SCF
less than 1 μm	76.8	0.08-0.12
1-2 μm	16.6	0.02-0.03
2-3 μm	4.7	0.005-0.008
3-4 μm	1.6	0.002-0.003
4-5 μm	0.2	0.0002-0.0003
5-6 μm	0.1	0.0001-0.0002
6-7 μm	0.0	-------------

Two-stage separation, final collector advanced design Aerodyne cyclone

This dust-containing gas will be at a pressure of approximately 10 atm and at a temperature of 1600°F. Approximately 50% of the dust will have been previously exposed to 2000°F temperature in the carbon burn-up cell or the sorbent regenerator. The balance of the ash will have experienced a maximum temperature of 1750°F in the fluid bed. These temperature maximums affect both the erosive properties of the solid dust particles and the corrosive nature of the gas stream. Three aspects of turbine operation are of special concern:

- Erosion of turbine rotor blades and stator vanes or nozzles by the entrained dust
- Corrosion of the blades by reaction of the hot sulfur- and oxygen-containing gas with the blade alloys, accelerated by sodium and potassium sulfates derived from the ash
- Deposition of material on the blade and vane surfaces, resulting in loss of efficiency of turbine stages.

Gas turbine operation on other than clean fuels -- natural gas or refined fuel oils -- has provided many examples of problems arising from erosion, corrosion, and blade fouling. It is important to keep in mind the differences in particle size distribution, physical and chemical nature of the ash (resulting from exposure to 1750 to 2000°F fluid bed temperatures rather than 2500 to 3000°F gas turbine combustor temperatures), and the smaller excess air ratios (10 to 1 instead of 50 to 1) while assessing the difficulty of efficiently operating gas turbines in the fluidized bed boiler power plant.

Erosion

Serious erosion of gas turbine rotor blades has been experienced in the operation of:

- Aircraft gas turbines which ingest large quantities of dust-laden air [67,91]

- Gas turbines combusting pulverized coal [5,86,98]
- Gas turbines used as power recovery units expanding hot gases from catalytic cracking operations [93]
- Gas turbines combusting and recovering energy from blast furnace gases [47].

Operating experience with turbines in these applications and laboratory studies of blade erosion [33,62] suggest several conclusions.

The majority of particles smaller than two micrometers in diameter can be expected to pass through the turbine without impacting on the blades. Those particles that do impact on the blades will have a polishing rather than an erosive effect. [62,67,93] However, particles larger than 10 micrometers in diameter are erosive. They impact on the blading with velocities and impact angles that are damaging [33,62]. These larger particles can be effectively removed by well-designed separators. No particles larger than six micrometers are expected in the gases leaving the carbon burn-up cell using advanced design Aerodyne separators. (Table 1-17)

Particulates entrained in the off-gases of the fluidized bed boiler have been subjected to lower temperature than those encountered by the ash in conventional gas turbine combustors. Only 50 percent of the ash has seen temperatures approaching 2000°F. This temperature is below the initial softening temperature of the vast majority of U.S. coals [1]. These particles have not fused to hard, potentially erosive glassy particles. Electron micrographs of ash particles from the British Coal Utilization Research Association's fluid bed pilot units showed that the larger particles were platelets of low mechanical strength. Although this bed operated at 1500°F, increasing bed temperature to 1750°F would not be expected to change materially the character of the ash, since the initial ash deformation and softening temperatures are not approached. Ash from the fluidized bed boiler is expected to be less erosive than the fused ash resulting from conventional combustion processes. The ash should also be less erosive than the catalyst particle carry-over responsible for erosion in oil refinery gas turbine applications. In this application, the use of three stages of mechanical dust collection resulted in turbine operating life in excess of four years [93].

Finally, dust loadings of 0.01 grains per standard cubic foot are projected to give acceptable turbine life, based on testing of the Locomotive Development Committee's coal-burning gas turbine. This unit was operated for over 4000 hours at Dunkirk, N.Y. and later for 2000 hours at the Bureau of Mines, Morgantown, W.Va. [86]. Tests were carried out using combusted pulverized coal having the particle size distribution and concentrations shown in Table 1-18.

TABLE 1-18

ASH PARTICLE SIZE AND CONCENTRATIONS USED TO ESTABLISH PERMISSIBLE LOADINGS IN BUREAU OF MINES COAL-FIRED GAS TURBINE RESEARCH PROJECT

PARTICLE SIZE (MICROMETERS)	WT. PERCENT	CONCENTRATION (GRAINS PER SCF)
less than 2	------	-----
2-5 μm	73.5%	0.088
5-10 μm	16.8	0.020
10-20 μm	7.3	0.009
20 μm	2.4	0.003
	100.0	0.120

Under a loading of 0.12 grains per SCF, the useful life of rotor blading was projected, based on 2000 hours of testing, to be 20,000 to 30,000 hours. Stator vane useful life was projected to be 5,000 to 7,000 hours. It was concluded that a factor of ten reduction in dust loading to 0.01 grains per standard cubic foot should give acceptable blade and vane life.

The Bureau of Mines did carry out erosion tests at their projected acceptable ash loading of 0.01 grain per standard cubic foot. Unfortunately, these tests were carried out at gas velocities unrealistically high for utility gas turbines and were made with 94 weight percent of the ash larger than 5 micrometers in diameter. A 1000-hour test under these conditions resulted in erosion of both the front and back surfaces of the blades [98].

The Bureau's original projection of an acceptable dust loading of 0.01 grains per standard cubic foot, especially in view of the less erosive nature of fluidized bed ash, is considered valid for the fluidized bed boiler system.

Erosion damage should not be a serious problem in turbines used with the fluidized bed boiler system provided that the dust loadings anticipated in Table 1-17 can be maintained. The concentration of particles larger than two micrometers in diameter should be between 0.008 and 0.012 grains per standard cubic foot. No particles larger than ten micrometers in diameter are expected to pass the collection system.

Careful design of the turbine to avoid concentrating the ash through the action of secondary flows within the turbine is essential to avoid erosion damage. Entrainment of ash in the wake formed at the stator vane trailing edges, resulting in ash concentration near the axis of the machine, caused erosion of rotor blade roots at their leading edges in both U. S. Bureau of Mines and Australian tests [5,86].

As a result of operating experience, the Bureau of Mines had their coal-fired turbine redesigned to incorporate:

- An annulus immediately following the first-stage stator vanes formed by omitting the first row rotor blading, to provide space for centrifuging ash particles to the outer wall of the turbine

- Thickened trailing edges of the rotor blading to encourage radially outward secondary flow in the rotor blade wakes
- Thinned trailing edges of stator vanes, to inhibit radially inward directed flow in the wakes of these vanes
- Added vanes to the stator rows, to minimize flow separation on the faces of the vanes. These flow separations had the effect of contributing ash to the secondary flow directed toward the rotor axis.
- Stepped side walls so that the tips of the stator vanes ended below the rotor blade platforms and tips of rotor blades extended above the stator vane roots
- Titanium carbide wear rings, to armor the rotor blade and stator vane roots against wear.

The Australian turbine designers redesigned their turbine to overcome erosion problems by:

- Adding a boundary layer fence, mid-span on the trailing edge of the stator vanes, to break up secondary flows
- Extending stator vane tips below the platform of adjacent rotor blades, so that ash leaving the stator blade tips would be diluted with cooling air and swept under the rotor blade root platforms
- Increasing turbine diameter to lower the velocity of gases through the turbine. Because erosion damage is a strong function of velocity (damage proportional to velocity raised to a power greater than two), lowering velocity was considered the most important step taken to overcome erosion problems.

Corrosion

Accelerated "hot" corrosion of gas turbine blades results from the formation on the blade surfaces of liquid films which react to remove the protective oxide scale and to raise the sulfur activity in contact with the blade alloy to the point where chromium sulfide can form. Eventually the local level of chromium in the alloy drops to a level that permits the formation of low-melting nickel sulfides. The blade loses mechanical integrity and is opened to catastrophic oxidation. [11,24,36,72,83].

Sodium sulfate liquid films, formed by the reaction on blade surfaces of sulfur dioxide-trioxide mixtures and sodium chloride vapors, cause hot corrosion attack in turbines operating with inlet gas temperatures above 1550°F (840°C) [23]. (The melting point of Na_2SO_4 is 1623°F, 884°C.) In the presence of gaseous potassium compounds, reaction on blade surfaces to liquid films could take place at temperatures as low as 930°F (500°C) [9].

The fluidized bed contains both sodium and potassium compounds which will exert small, but potentially harmful vapor pressures at the bed operating temperature.

Operation of the bed at 1750-1800°F lowers the vapor pressures exerted by the compounds approximately three orders of magnitude below those expected in conventional boiler furnaces. In addition, the fluid bed provides a large surface area of alumino-silicate ash and long residence times to encourage in-bed reactions. These reactions should act to lock up sodium and potassium in solid compounds, reducing the quantity of corrosive material leaving the bed. Detailed studies of the reactions between the bed materials and volatile sodium and potassium compounds are needed to provide quantitative data for design. It may be possible to eliminate the danger of hot corrosion and fouling if the actual partial pressures of the harmful condensable vapors can be reduced through gas-solid reactions below those allowing condensation in the turbine.

In the worst case, limitation of the temperature of the inlet gas to the turbine may be used to prevent liquid film formation on turbine hardware. Reduction of gas inlet temperature reduces the efficiency of the turbine, which in turn lowers the overall plant efficiency. Should it be necessary to limit the turbine inlet temperature to 1400°F, the turbine efficiency would drop from approximately 28 to 25%, resulting in about a one percent reduction in overall plant efficiency.

Combining a carefully engineered fluid bed boiler and dust collection system with an industrial gas turbine employing sulfidation-resistant high-chromium alloys in its construction should allow operation of the turbine without serious corrosion problems.

INDUSTRIAL BOILER APPLICATION

Better methods are needed for combusting the vast reserve of coal to meet the ever-increasing energy demand in the United States. Fluidized bed combustion stands out as a technology with the potential to combust different grades of coal efficiently and simultaneously to conform to increasingly stringent air pollution control regulations. Pollution from industrial boilers is not to be disregarded, especially that of SO_x. As a whole, industrial boilers emit about 14% of the total SO_x emissions and 11% of total NO_x emissions from power/steam generating facilities.

Industrial boilers are generally considered nonprofit equipment for which the desirable characteristics are high availability, low maintenance, and economy. Therefore, an industrial fluidized bed boiler that can be shop-fabricated and transported by rail is the most promising. The preliminary design of an industrial fluidized bed boiler was begun with these characteristics in mind. Different conceptual designs were evaluated in conjunction with the Erie City Energy Division of Zurn Industries, a Westinghouse subcontractor. From this preliminary evaluation, two systems were selected for detailed design and cost analysis: one-through dry solid and wet scrubbing. Costs of other air pollution control alternatives were also derived and compared with these two systems.

Design Basis

System Specifications

The system specifications for the industrial fluidized bed boiler are summarized in the first specification list. The plant capacity and steam conditions were selected on the basis of the market survey, which projects that 40% of new units in 1980 will be in this category. At the chosen capacity, the boiler remains shop-fabricable and railroad-transportable.

Fuel and Limestone Specifications

The ultimate analysis and proximate analysis of the Ohio Pittsburgh No. 8 seam coal are presented in the second specification list. The particle size distribution of feed coal as crushed from Koppers' reversible hammermill is shown in Figure 1-64. The ash residual after combustion in the fluidized beds was assumed to be 100% elutriated on the basis of experimental evidence from the U. S. Bureau of Mines laboratories at Morgantown. The particle size-fluidizing velocity relationship is presented in Figure 1-65.

Limestone No. 1359, with composition analysis as shown in Table 1-19, was chosen for the industrial fluid bed application.

Design

The designs for the fluidized bed combustor (FBC) and the carbon burn-up cell (CBC) are summarized in Table 1-20. The physical dimensions of the boiler were kept within the railroad transportation limitations, which are roughly 13 feet wide, 16 feet high, and 40 feet long for most eastern United States locations.

SYSTEM SPECIFICATIONS FOR INDUSTRIAL BOILER

PLANT CAPACITY & STEAM CONDITIONS

250,000 lb/hr steam at 600 psig and 750°F

CYCLE CONDITIONS

Gas	-- air inlet temperature	80°F
	stack gas temperature	350°F (once-through)
		>130°F (wet scrubbing)
Water	-- water inlet temperature to economizer	250°F
	water outlet temperature from economizer	350°F

PARTICULATE REMOVAL REQUIREMENT

Flue gas particulate loading	0.01-0.05 gr/SCF

SULFUR REMOVAL REQUIREMENT

>90%

TURN-DOWN REQUIREMENT

Turn-down ratio	$\geq 4:1$
Dynamic response	$\geq 5\%$/min

OHIO PITTSBURGH NO. 8 SEAM COAL

(Source of data: USBM, Pittsburgh, Pa.)

SAMPLE: Run of mine - as received

PROXIMATE ANALYSIS (wt %):

Moisture	3.3
Volatile matter	39.5
Fixed carbon	48.7
Ash	8.5
	100.0

ULTIMATE ANALYSIS (wt %):
(includes moisture)

C	71.2	
H	5.4	
O	9.3	
N	1.3	
S	4.3	(∿60% organic & ∿40% pyritic)
Ash	8.5	
	100.0	

GROSS HEATING VALUE: 13000 Btu/lb

NET HEATING VALUE: 12500 Btu/lb

ASH ANALYSIS (wt %):

SiO_2	45.3
Al_2O_3	21.2
Fe_2O_3	27.3
TiO_2	1.0
P_2O_5	0.11
CaO	1.9
MgO	0.6
Na_2O	0.2
K_2O	1.8
SO_3	0.7
	100.1

FUSIBILITY OF ASH:

Initial deformation temperature	-- 2080°F
Softening temperature	-- 2230°F
Fluid temperature	-- 2420°F

PARTICLE DENSITY:

Coal -- ∿1.4 gm/cc
Ash -- ∿2.8 gm/cc

GRINDABILITY (Hardgrove): 50-60

FREE SWELLING INDEX: 5-5.5

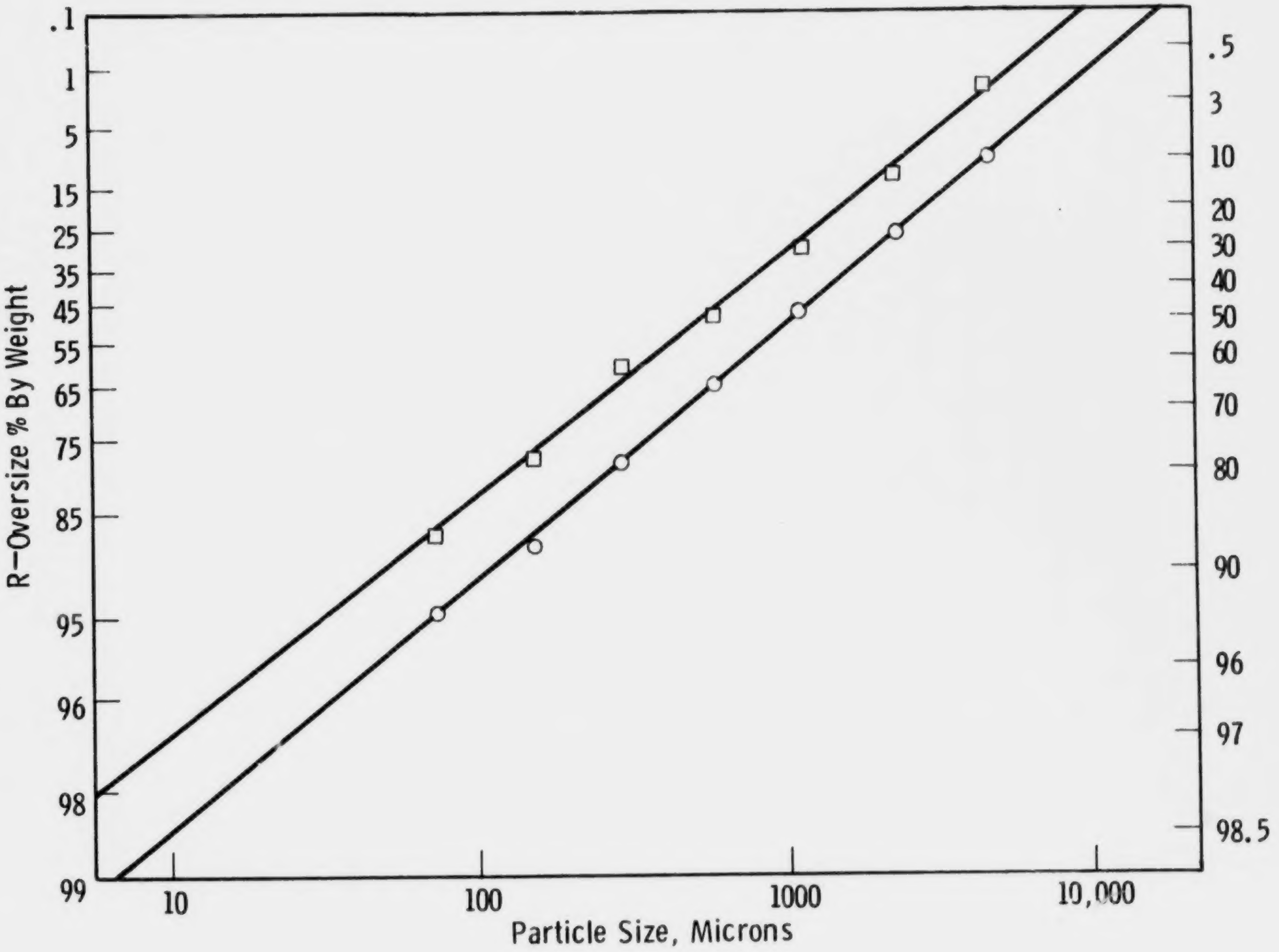


Fig. 1-64–Size consists of coal crushed by Koppers Reversible Hammermill

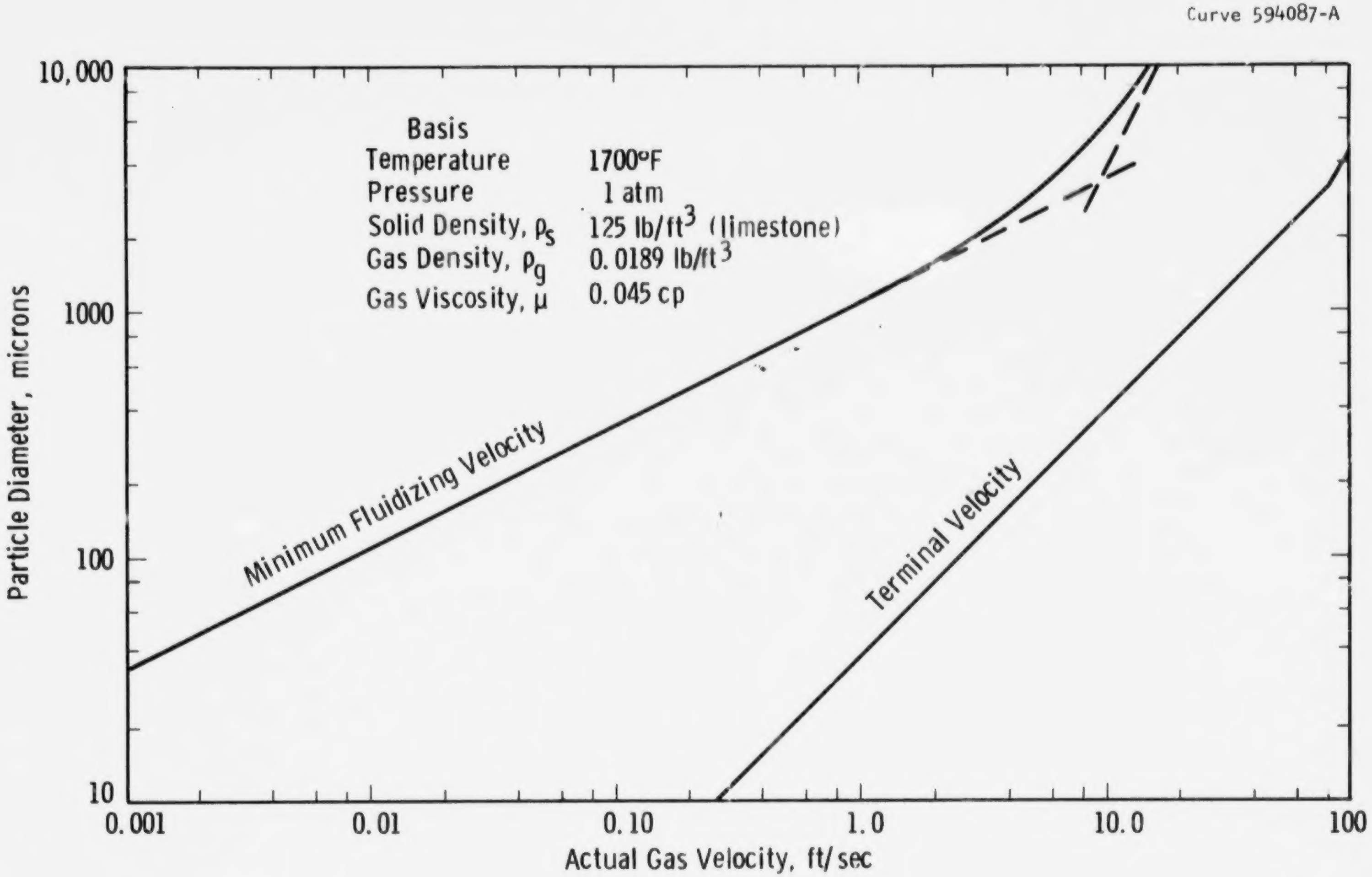


Fig. 1-65—Fluidizing velocity and terminal velocity for 1 atm system

TABLE 1-19

ANALYSIS OF LIMESTONE NO. 1359

COMPONENT	WT % AS RECEIVED[a]
SiO_2	0.85
Al_2O_3	0.30
Fe_2O_3	0.17
MgO	1.07
CaO	97.
TiO_2	< 0.05
SrO	0.07
Na_2O	< 0.02
K_2O	< 0.1
MnO_2	< 0.05

[a] R. W. Coutant, J. S. McNulty, R. E. Barrett, J. J. Carson, R. Fischer, and E. H. Lougher; "Investigation of the Reactivity of Limestone and Dolomite for Capturing SO_2 from Flue Gas," August 1968 (Battelle Memorial Institute).

TABLE 1-20 DESIGNS FOR INDUSTRIAL FLUIDIZED BED BOILER

	FLUIDIZED BED COMBUSTOR (FBC)	CARBON BURN-UP CELL (CBC)
Bed Temperature, ^{o}F	1650	1900-2000
Superficial fluidizing velocity, ft/sec	12.5	10
Bed depth (expanded), in	30	30
Excess air, %	10	50
Carbon burn-up efficiency, %	87	90
Bed material	Limestone No. 1359	Limestone No. 1359
Gas side heat transfer coefficient, BTU/ft^2-hr-°F	50 in the bed 40 in stop-splashing screen	50 in the bed
Sulfur removal schemes	Wet scrubbing or once-through	
Stone	Limestone No. 1359	
Ca/S ratio	1.2/6	

The size distribution of the particles elutriated from the FBC and CBC, shown in Figure 1-66, was also projected on the basis of experimental data accumulated in the laboratories of the National Coal Board (NCB) and the British Coal Utilization Research Association (BCURA) in England, and of the Office of Air Program (OAP) contractors in the United States. This data correlation was discussed on pages 119-131.

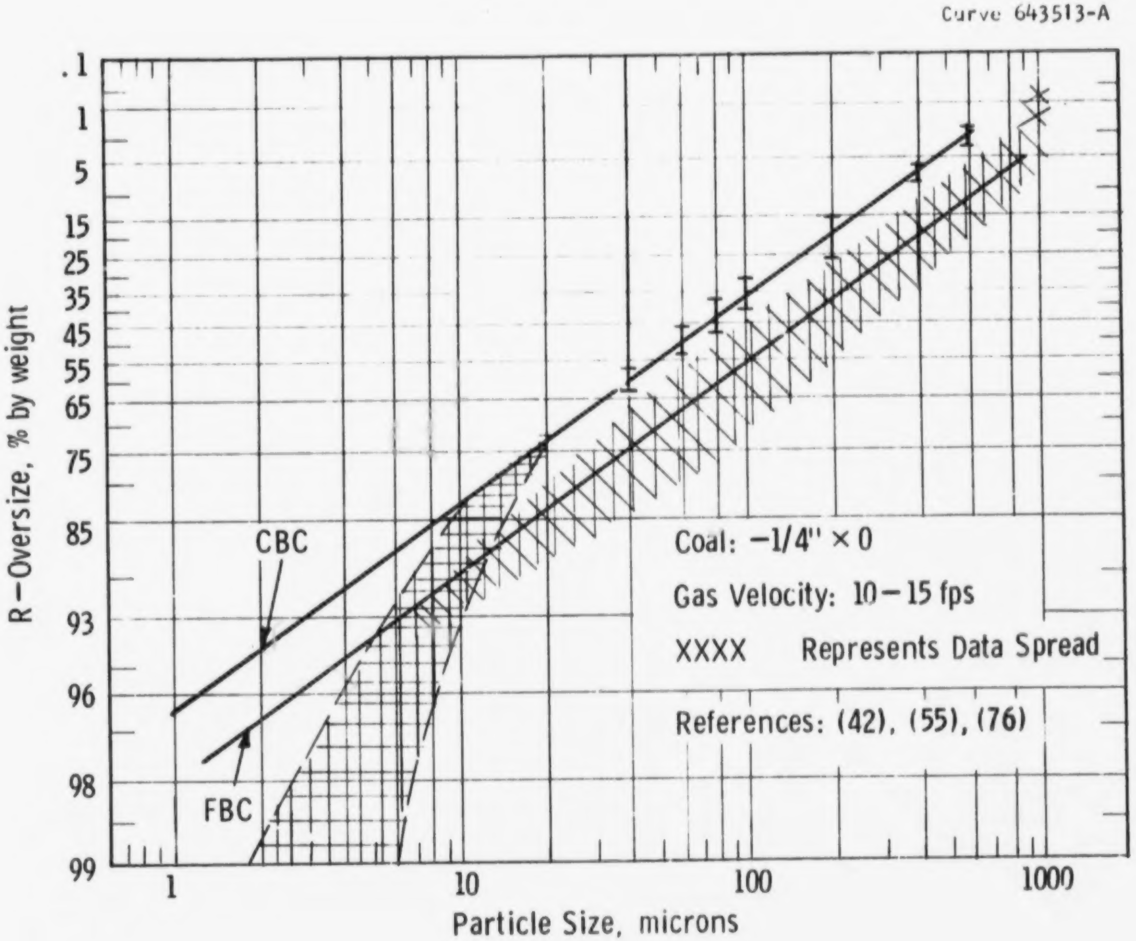


Fig. 1-66 – Projected particle size distribution for material elutriated from fluidized bed combustor (FBC) & carbon burn-up cell (CBC)

Boiler System Design

The overall boiler system design is outlined here. The flow sheets and energy/mass balances are summarized on pages 174-177 . Different boiler design concepts are discussed on pages 177-183 , and the final design selected is described on pages 183-191 . Other auxiliary systems include the coal and limestone feeding system, particulate removal system, and air pollution control systems; these are presented on pages 191-195 , and 196-198 respectively. The detailed design is presented in Appendix D.

Boiler System Schematic

The overall flow sheets for the two chosen systems -- once-through dry solid and wet scrubbing -- were simplified and presented in Figures 1-67 and 1-68. The detailed schematics are included in Appendix D.

For the once-through dry solid system, six times stoichiometric limestone is fed to the primary bed and calcined. Although this will not be favorable in industrial applications, it is included in this analysis for comparison for two reasons. First, the waste solid rate is comparable to that of the wet scrubbing system as shown in Table 1-21. Second, for the once-through limestone application large quantities of stone, on the order of six times stoichiometric limestone, are required for 90% SO_2 removal at design conditions. Ninety percent of SO_2 will be removed in the bed; this represents $\sim$16% stone utilization. The limestone is used once-through without regeneration. Carbon-rich ash is recycled back to the carbon burn-up cell. The flue gas passes through a superheater, convection zone, economizer, and electrostatic precipitator with an efficiency of > 99.5%. The particulate loading in flue gas is therefore expected to be below 0.02 gr/ft^3.

Dwg. 2957A32

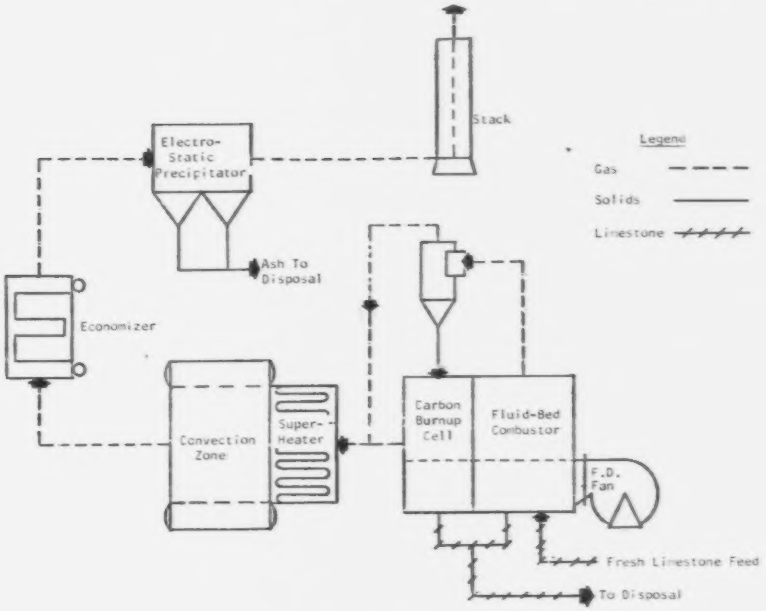


Fig. 1-67 – System Schematic - Once-Through Scheme

Dwg. 2957A31

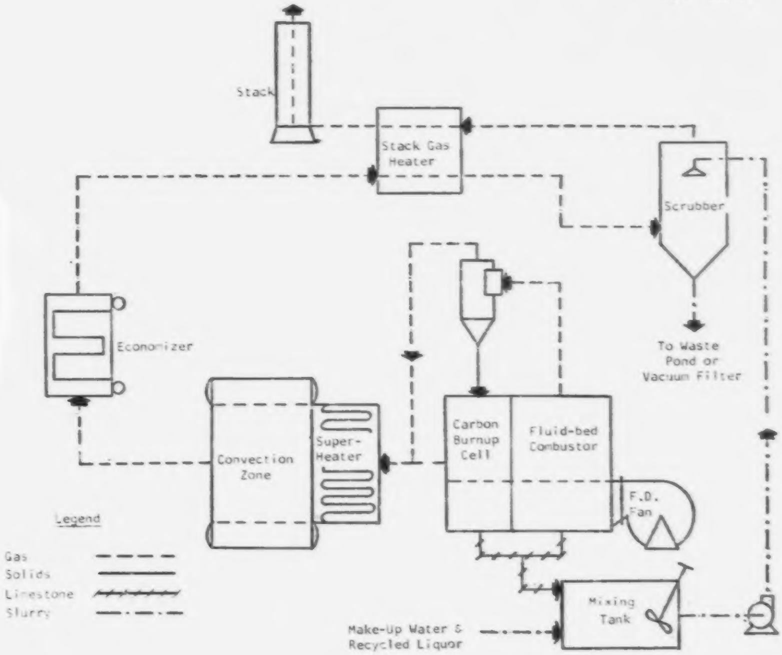


Fig. 1-68 – System Schematic - Wet Scrubbing Scheme

TABLE 1-21

COMPARISON OF BOILER PERFORMANCE

	DRY SOLID	WET SCRUBBING
COAL FEED RATE (lb/hr)	36,650	25,900
LIMESTONE FEED RATE (lb/hr)	21,450	4,200
WASTE SOLID RATE (lb/hr)	17,300	16,050
BOILER EFFICIENCY (%)	83.42	85.92
TOTAL LOSS (%)	16.58	14.08
Loss due to calcination and SO_2 absorption, %	2.74	0.56
Loss due to water and hydrogen in fuel, %	3.80	3.83
Loss due to unburned carbon, %	1.93	2.05
Loss due to radiation and unaccounted for, %	1.50	1.50
Loss due to evaporation in wet scrubber, %	-	4.84
Loss due to flue gas sensible heat, %	6.61(flue gas temperature = 350°F)	1.30(flue gas temperature = 132°F)

For the wet scrubbing system, 1.2 times stoichiometric limestone is fed to the primary bed and calcined. The calcined limestone is then mixed with water to form slurry and pumped to the wet scrubber. The fluidized bed boiler design is essentially the same as that in the once-through system. The only difference is that the flue gas, instead of passing through the electrostatic precipitator, goes to the wet scrubber for SO_2 and particulate removal. The stack gas heater is omitted in the present design, so the stack gas temperature is $\sim$130°F.

Energy and Material Balances

Energy and material balances for both once-through dry solid and wet scrubbing systems are shown in Figures 1-69 and 1-70. The comparison of the boiler efficiencies of these two systems is summarized in Table 1-21. Heat generation from SO_2-CaO reaction is assumed to be 3.0×10^6 BTU/ton $CaSO_4$ produced; heat absorption from calcination of $CaCO_3$ is assumed to be 1.5×10^6 BTU/ton $CaCO_3$ calcined.

The wet scrubbing system has higher boiler efficiency because the stack gas temperature is 132°F, compared to 350°F in the dry solid system. An additional efficiency loss of about 5% in the wet scrubbing system is expected if the flue gas is to be heated from 132°F to 350°F.

Industrial Boiler Design Concepts

A shop-fabricated and railroad-transportable industrial boiler is highly desirable. Rectangular beds are therefore more suitable for industrial boilers, for they make the best use of the available shipping clearance. A horizontal configuration in which fluidized beds are placed next to each other has clear advantages over the vertical configuration. First, during start-up or load change, solid circulation from an operating bed can be used to light off the idled beds and makes it unnecessary to provide large ignitors for individual beds. Second, maintenance is easier, and fewer support structures are needed.

Submerged heat transfer surface as vertical or horizontal tube design were both considered in the original evaluation. The vertical

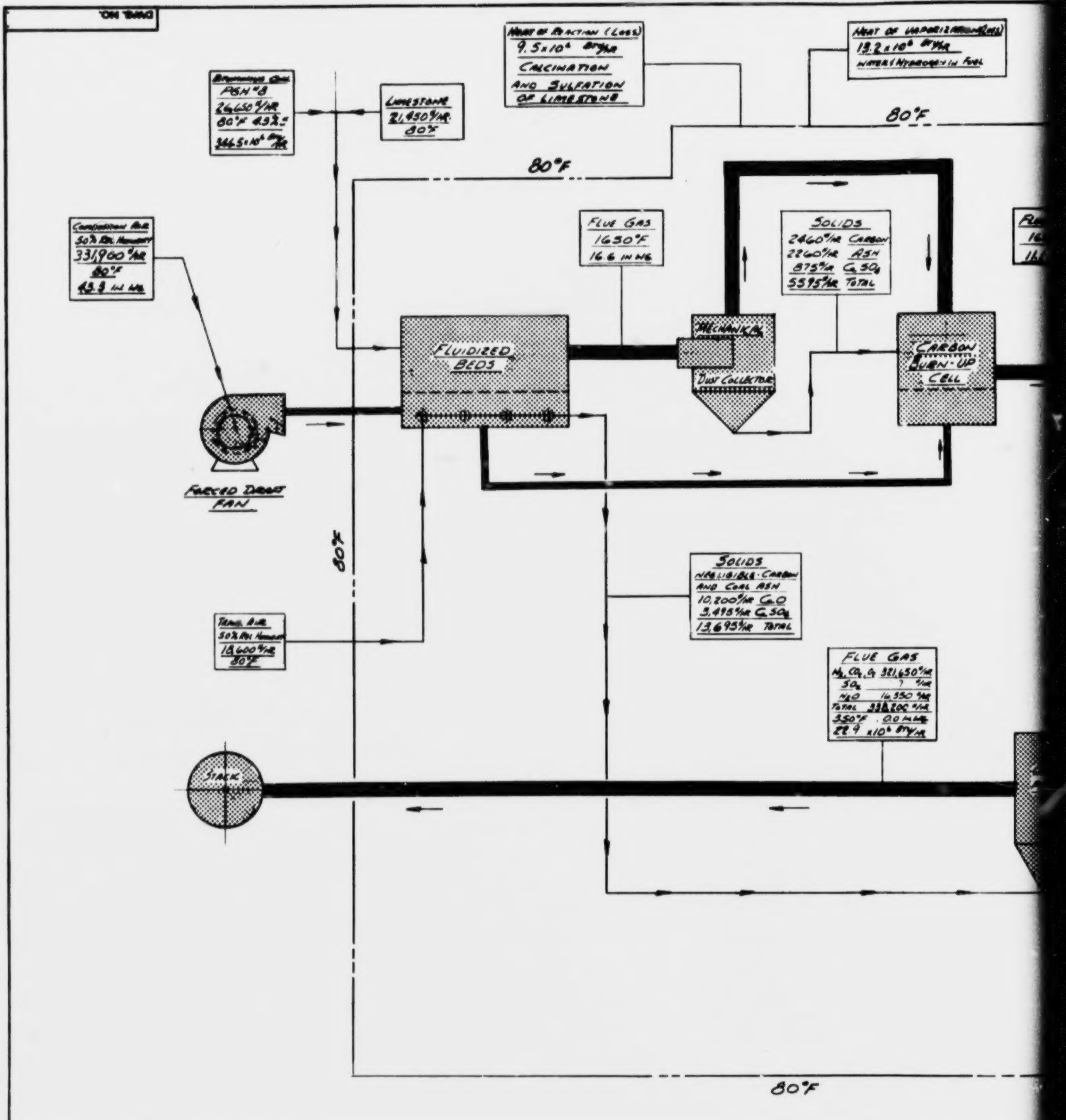
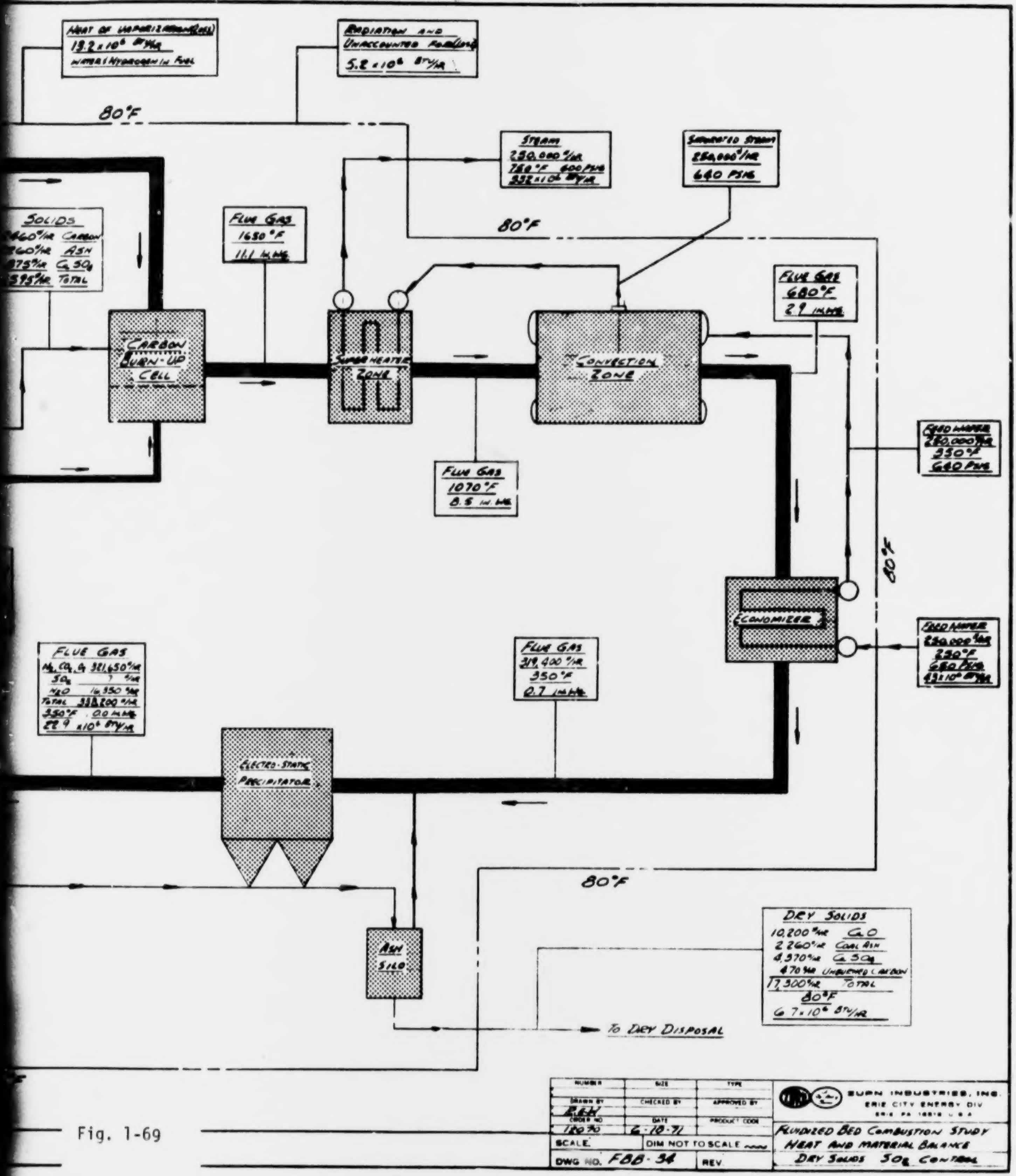
DWG. NO.
BITUMINOUS COAL
PSN #8
24,650 #/HR
80°F 4.3% S
346.5×10⁶ BTU/HR
LIMESTONE
21,450 #/HR
80°F
HEAT OF REACTION (LOSS)
9.5×10⁶ BTU/HR
CALCINATION
AND SULFATION
OF LIMESTONE
HEAT OF VAPORIZATION (LOSS)
19.2×10⁶ BTU/HR
WATER (HYDROGEN IN FUEL
80°F
80°F
COMBUSTION AIR
50% REL HUMIDITY
331,900 #/HR
80°F
43.3 IN WG
FLUE GAS
1650°F
16.6 IN WG
SOLIDS
2460 #/HR CARBON
2260 #/HR ASH
875 #/HR CaSO4
5595 #/HR TOTAL
FLUIDIZED BEDS
MECHANICAL
DUST COLLECTOR
CARBON BURN-UP CELL
FORCED DRAFT FAN
80°F
SOLIDS
NEGLIGIBLE CARBON
AND COAL ASH
10,200 #/HR CaO
3,495 #/HR CaSO4
13,695 #/HR TOTAL
TRANS. AIR
50% REL HUMIDITY
18,600 #/HR
80°F
FLUE GAS
N2, CO2, O2 321,650 #/HR
SO2 ? #/HR
H2O 16,350 #/HR
TOTAL 338,200 #/HR
350°F 0.0 IN WG
22.9 ×10⁶ BTU/HR
STACK
80°F

Fig. 1-69


HEAT OF VAPORIZATION (LOSS)
19.2 x 10⁶ BTU/HR
WATER (HYDROGEN IN FUEL
RADIATION AND
UNACCOUNTED FOR (LOSS)
5.8 x 10⁶ BTU/HR
80°F
STEAM
250,000 #/HR
750°F 600 PSIG
332 x 10⁶ BTU/HR
SATURATED STEAM
250,000#/HR
640 PSIG
SOLIDS
460#/HR CARBON
260#/HR ASH
875#/HR CaSO4
595#/HR TOTAL
FLUE GAS
1650°F
11.1 IN.WG.
80°F
FLUE GAS
680°F
2.9 IN.WG.
CARBON
BURN-UP
CELL
SUPERHEATER
ZONE
CONVECTION
ZONE
FEED WATER
250,000#/HR
350°F
640 PSIG
FLUE GAS
1070°F
8.5 IN.WG.
80°F
ECONOMIZER
FEED WATER
250,000#/HR
250°F
680 PSIG
43 x 10⁶ BTU/HR
FLUE GAS
N2, CO2, O2 321,650#/HR
SO2 7 #/HR
H2O 16,550 #/HR
TOTAL 338,200 #/HR
350°F .0.0 IN.WG.
22.9 x 10⁶ BTU/HR
FLUE GAS
319,400 #/HR
350°F
0.7 IN.WG.
ELECTRO-STATIC
PRECIPITATOR
80°F
DRY SOLIDS
10,200#/HR CaO
2 260#/HR COAL ASH
4,370#/HR CaSO4
470#/HR UNBURNED CARBON
17,300#/HR TOTAL
80°F
6.7 x 10⁶ BTU/HR
ASH
SILO
TO DRY DISPOSAL
NUMBER
SIZE
TYPE
DRAWN BY
REH
CHECKED BY
APPROVED BY
ORDER NO
12030
DATE
6-10-71
PRODUCT CODE
SCALE
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DWG NO. FBB-34
REV.
ZURN INDUSTRIES, INC.
ERIE CITY ENERGY DIV
ERIE PA 16512 U.S.A.
FLUIDIZED BED COMBUSTION STUDY
HEAT AND MATERIAL BALANCE
DRY SOLIDS SO2 CONTROL

Fig. 1-69

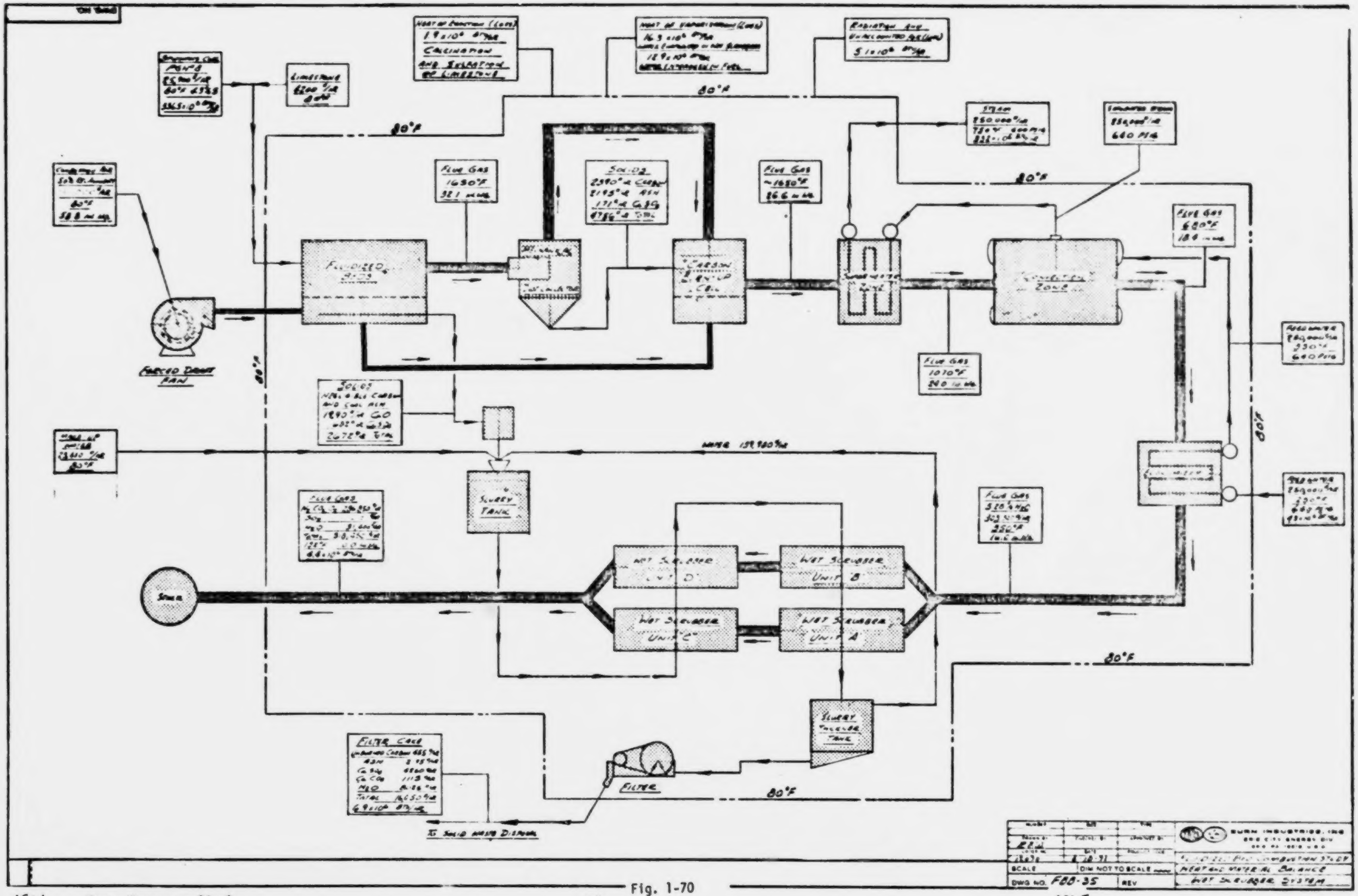

Fig. 1-70

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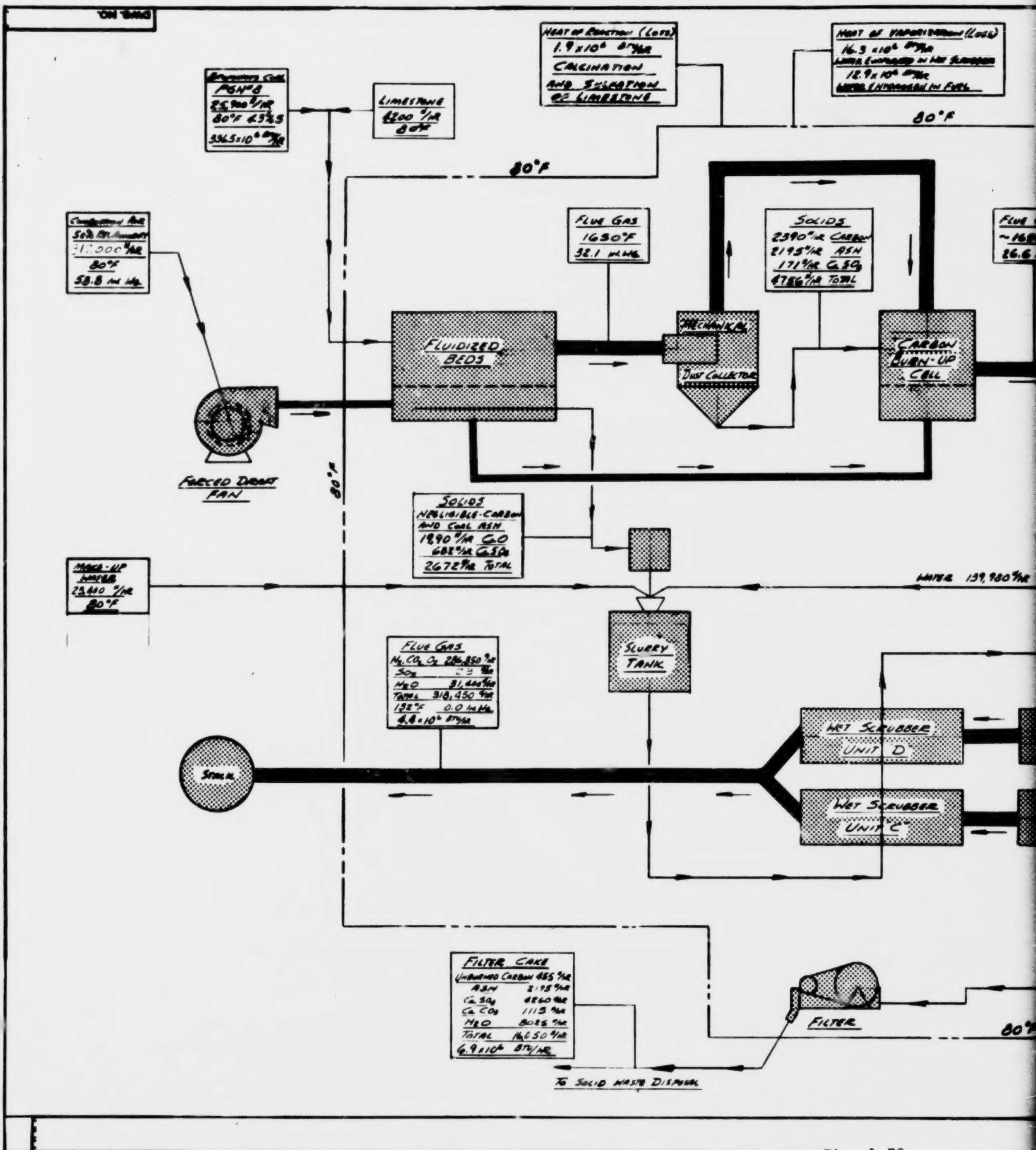

Fig. 1-70

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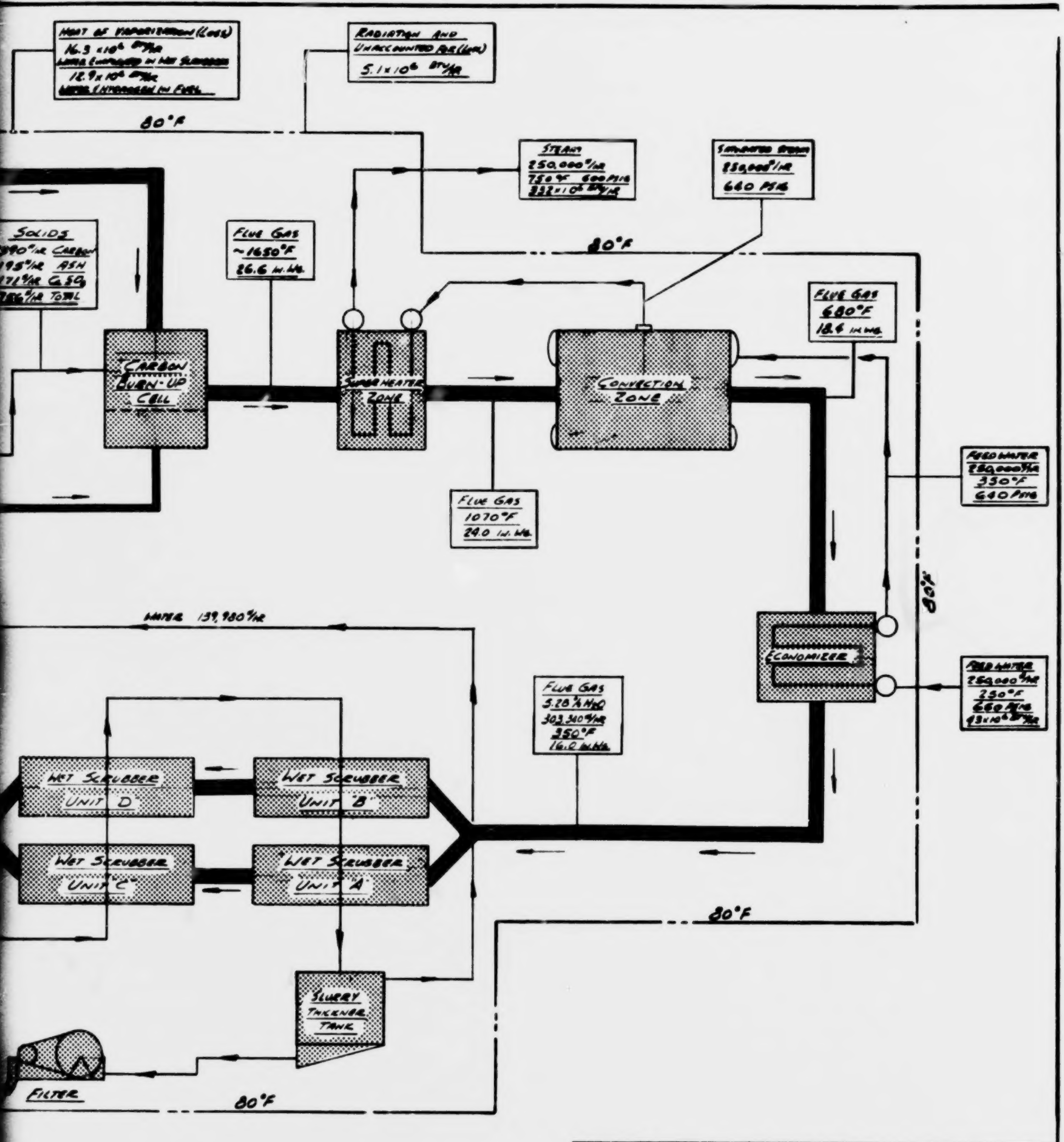

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ZURN INDUSTRIES, INC.
ERIE CITY ENERGY DIV.
ERIE PA 16512 U S A

FLUIDIZED BED COMBUSTION STUDY
HEAT AND MATERIAL BALANCE
WET SCRUBBER SYSTEM

Fig. 1-70

tube design used natural water circulation, while the horizontal tube design used forced water circulation. Vertical tube design was rejected when the preliminary cost analysis showed that its cost was about twice that of the horizontal tube design, primarily because of the large number of welds required. (See Appendix D).

The superheater in the selected design is conventional and is placed in the convection section. Putting a superheater in the carbon burn-up cell is considered inadequate for two reasons. First, to maintain carbon burn-up efficiency in the CBC, the bed temperature has to be kept constant, even during turn-down. With the boiler load decreased, the recycled carbon fines back to the CBC decrease. Therefore, the bed temperature will drop unless the bed depth is simultaneously decreased. The regulation of bed depth involves costly equipment and complex operation. In addition, for a load reduction to 25% of full load, the bed depth would be too shallow for efficient operation. Second, additional fresh coal can be fed into the CBC to keep the bed temperature constant without changing the bed depth. For a load reduction to 25% of full-load capacity, the steam flow in the superheater tubes is just about 25% of the full-load rate. Thus, the steam would be superheated to a much higher temperature and the tube metal temperature would be excessively high, requiring the use of expensive, high-grade alloy material.

Through the same kind of preliminary concept evaluation for each individual boiler component, a final boiler design was selected.

Selected Steam Generator Design

After evaluation of different boiler design concepts on the basis of presently available information, a boiler design with the overall isometric view shown in Figure 1-71 was chosen. The sectioned view is presented in Figure 1-72. The boiler comprises two separate shipping modules: the primary module (overall dimensions, 10 ft x 14 ft x 40 ft) consists of four fluidized beds with air plenum sectionalized into four sections, but not the bed interiors, the secondary module.

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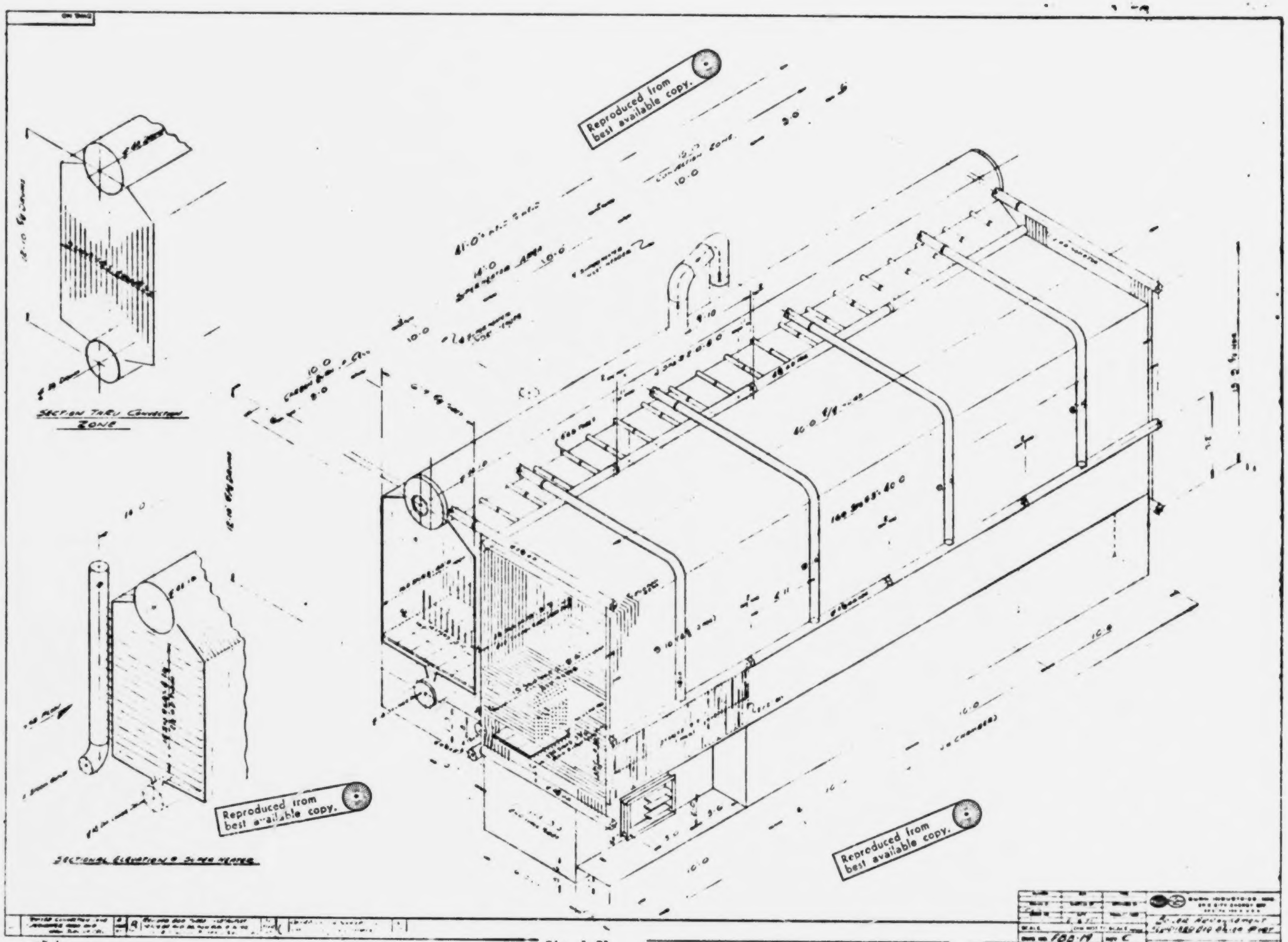
SECTION THRU CONVECTION ZONE
SECTIONAL ELEVATION @ SUPERHEATER
10'-0
10'-0
16'-0
Convection Zone
3'-0
100-14
C

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185.1

Fig. 1-71
185.2

185.3

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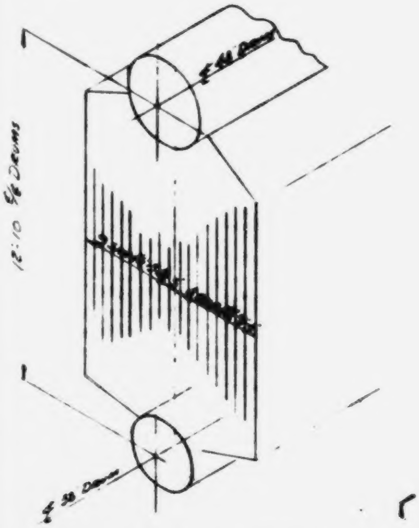
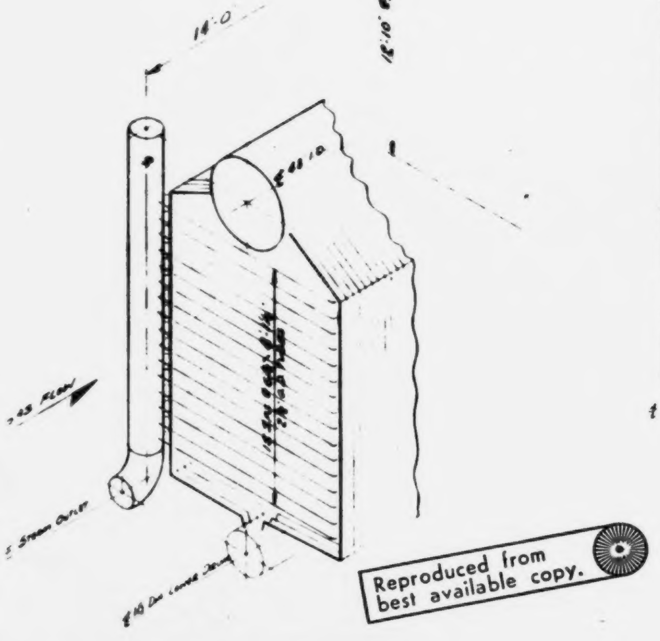


SECTION THRU CONVECTION ZONE



SECTIONAL ELEVATION @ SUPER HEATER

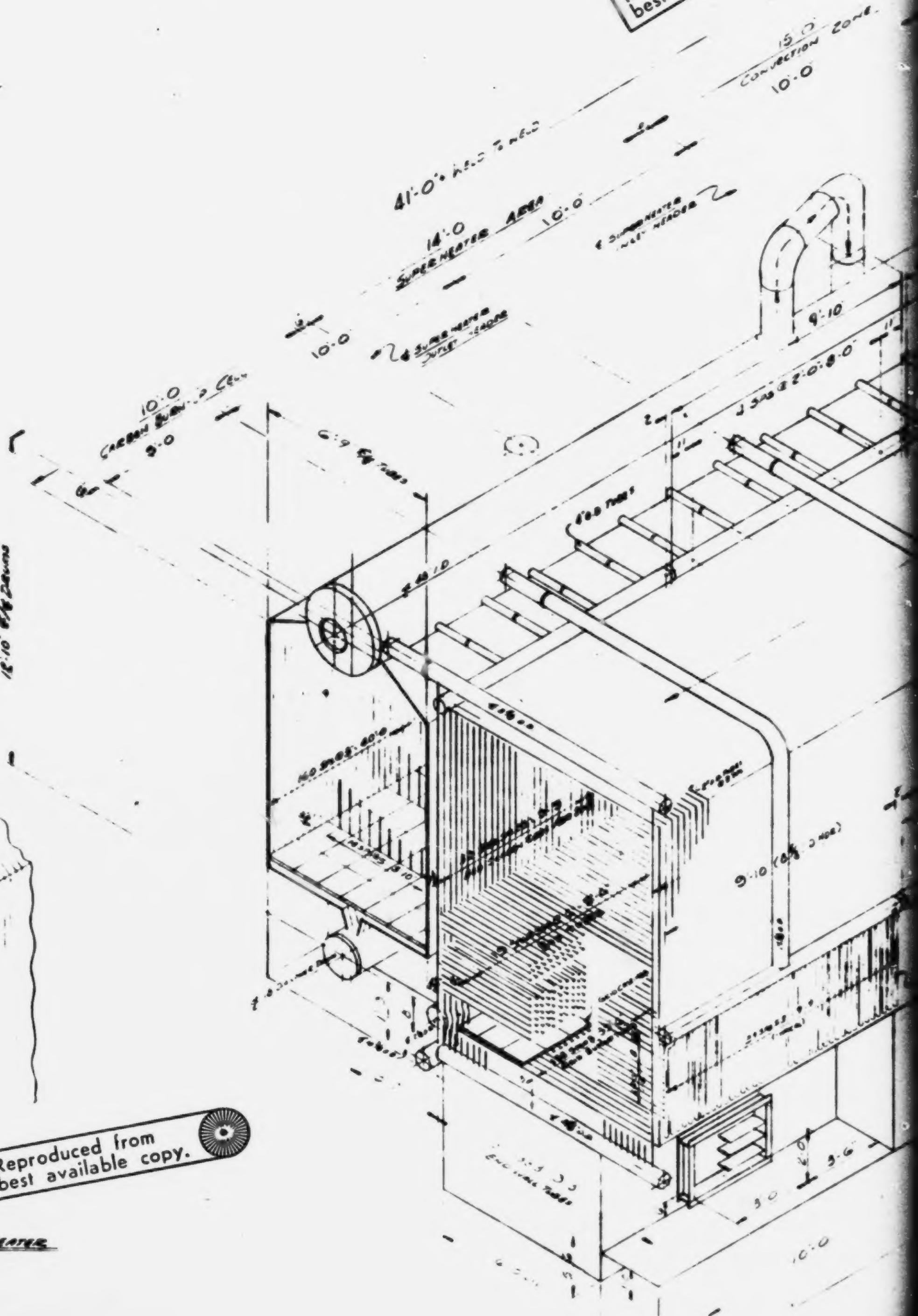


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Fig. 1-71

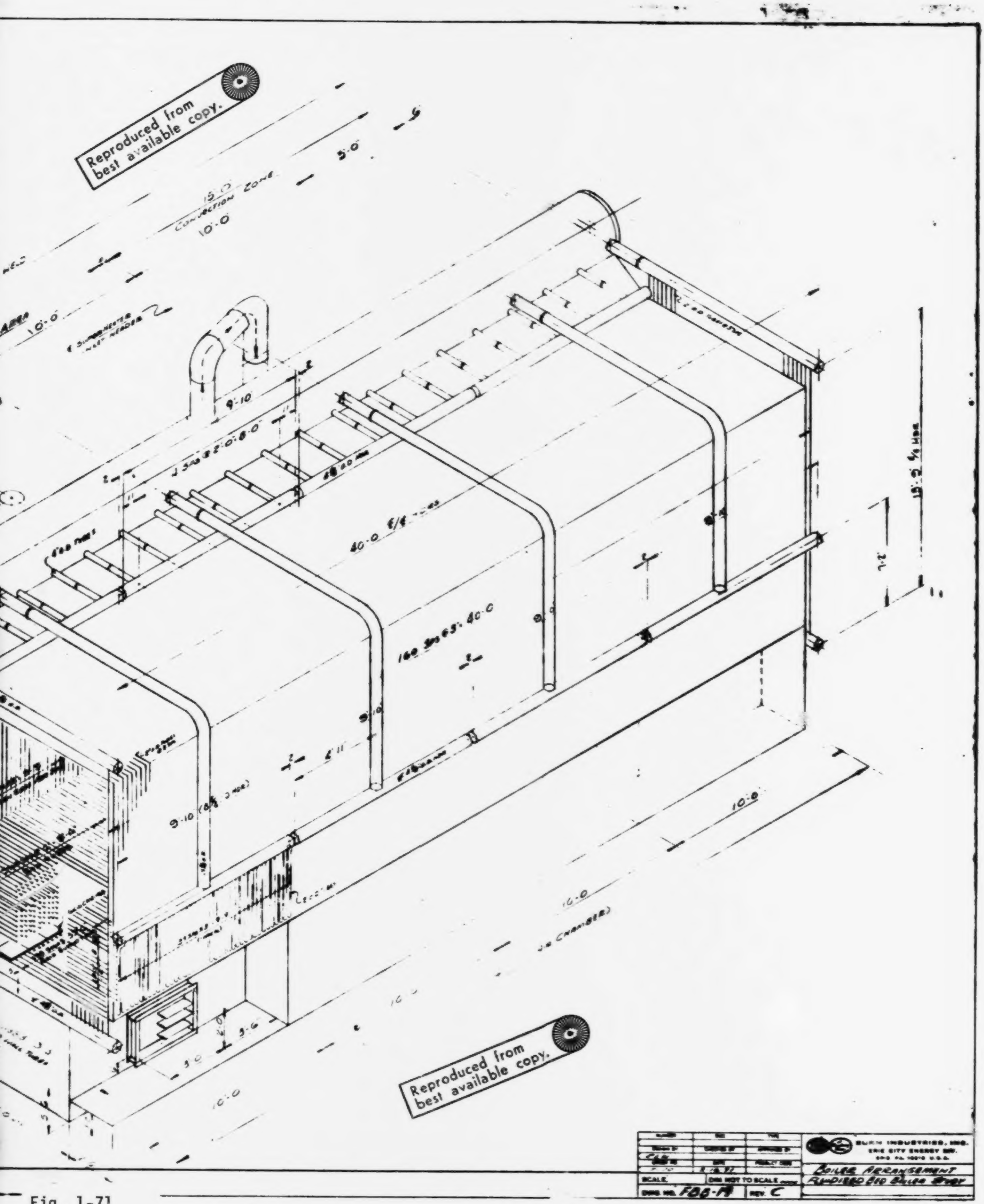

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Fig. 1-71
185.2

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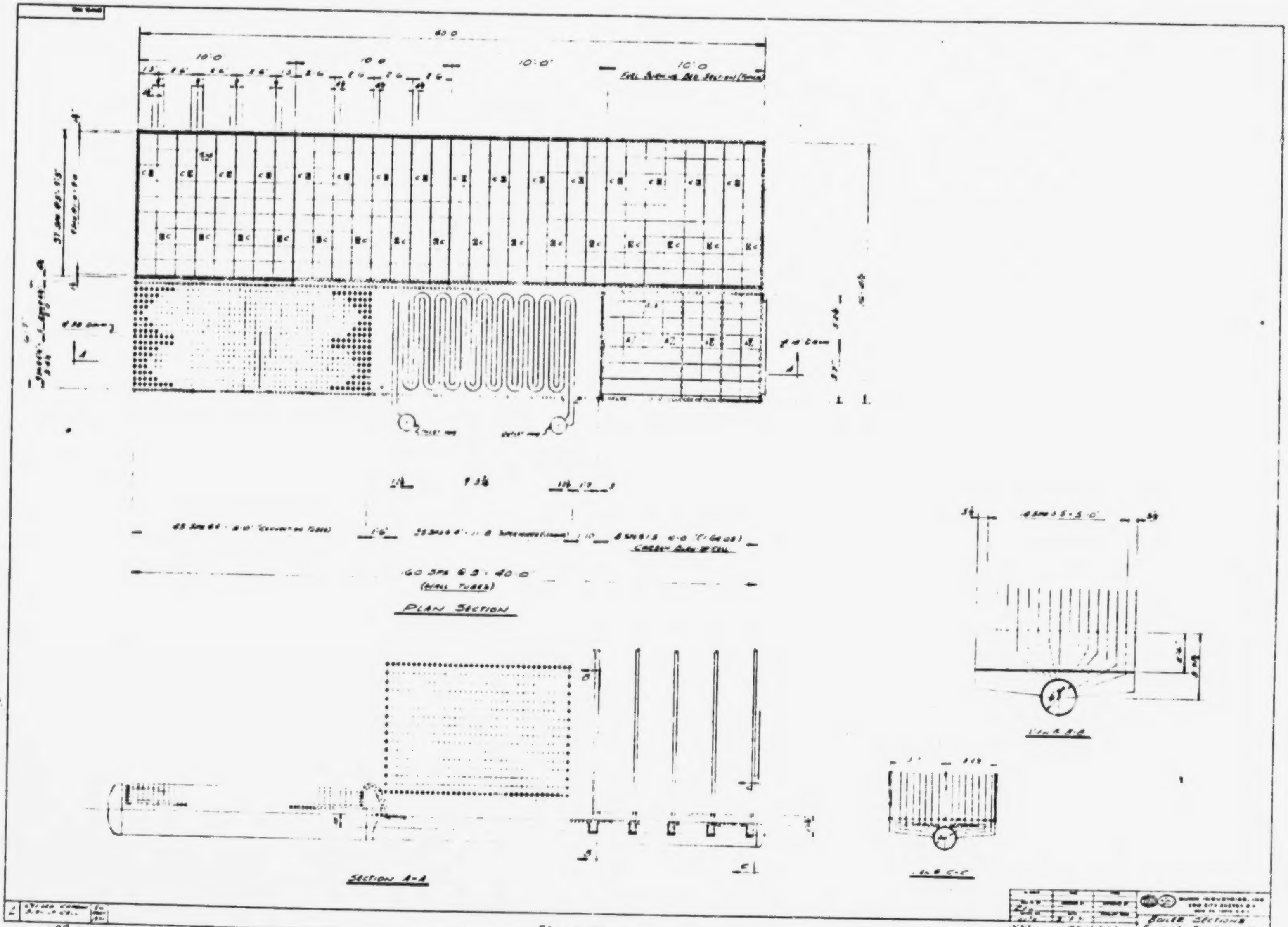

Fig. 1-72

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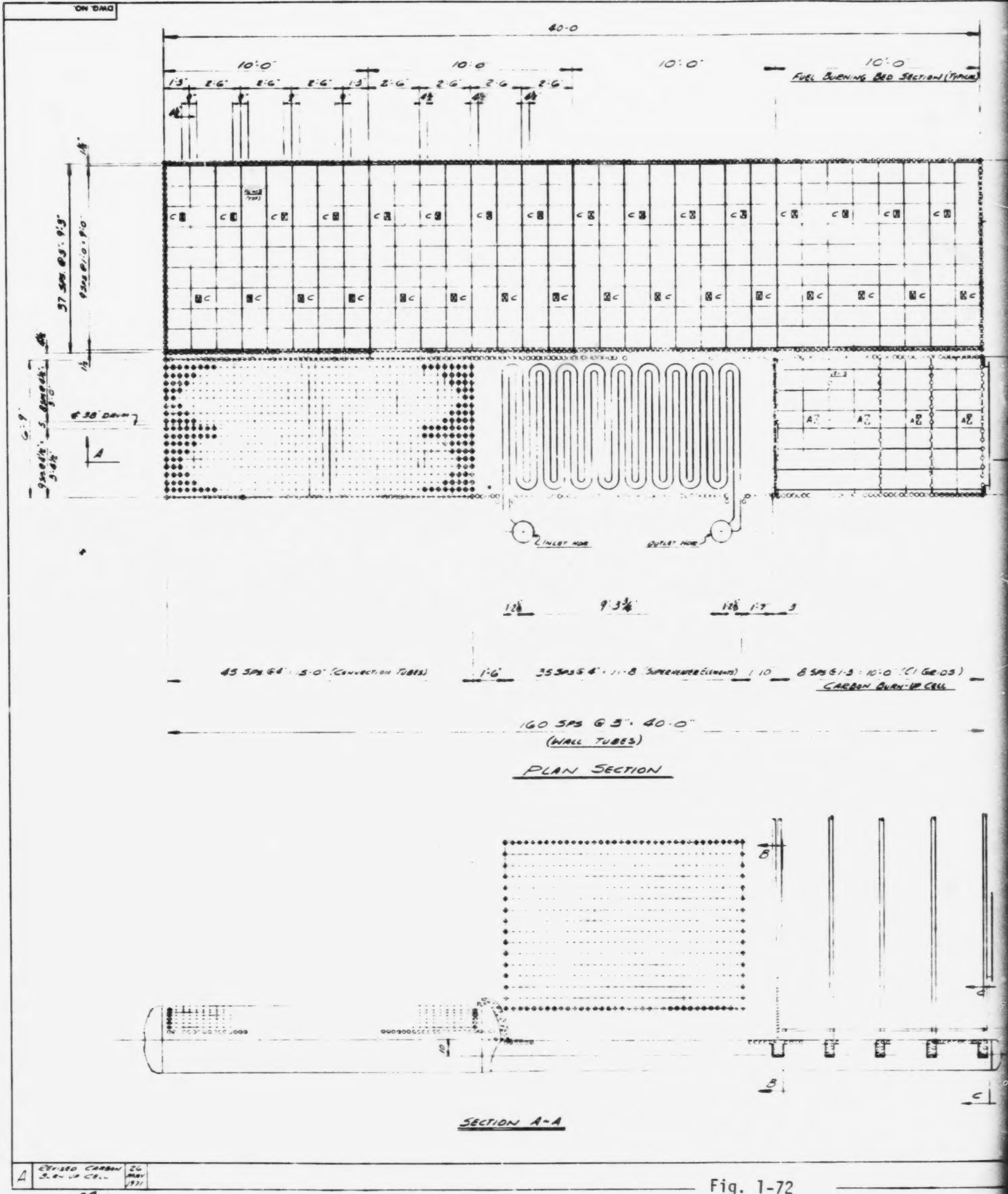
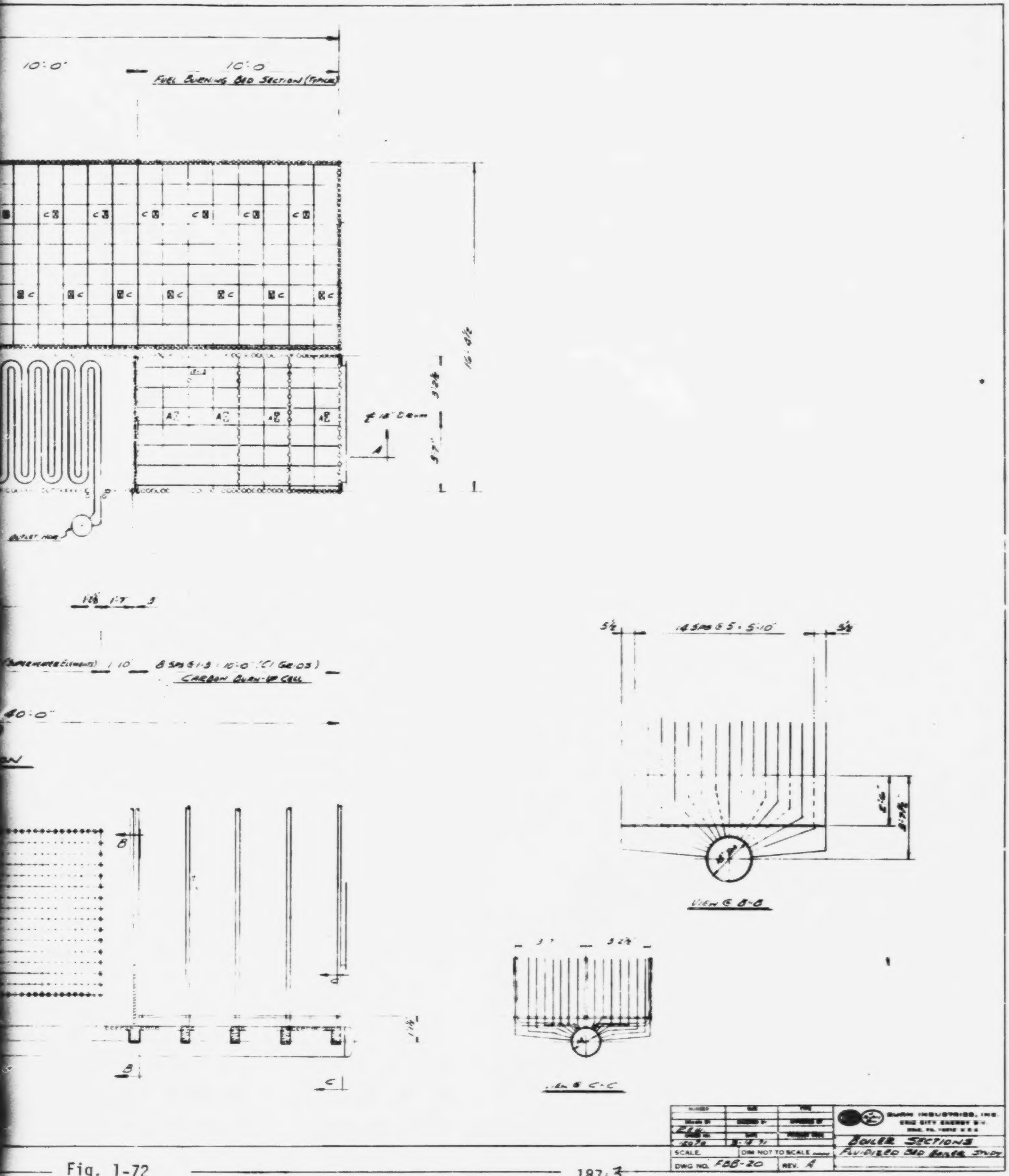


187-1

Fig. 1-72

187-2


10'-0"
10'-0"
FUEL BURNING BED SECTION (TYPICAL)
16'-8½
¢ 18" DRUM
3'2½
5'7"
A
OUTLET HDR.
1-8½ 1-7 3
(SUPERHEATER ELEMENTS) 1 10
8 SPS @ 1-3 = 10'-0" (C1 GRIDS)
CARBON BURN-UP CELL
40'-0"
B
B
C
C
5½
14 SPS @ 5 = 5-10
5½
VIEW @ B-B
3 7
3 2½
VIEW @ C-C
ZURN INDUSTRIES, INC.
ERIE CITY ENERGY DIV.
SCALE
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DWG NO. FBB-20
REV. A
BOILER SECTIONS
FLUIDIZED BED BOILER STUDY

Fig. 1-72

187.2

(overall dimensions, 8 ft x 15 ft x 40 ft) houses the carbon burn-up cell, superheater, and the convection pass. The carbon burn-up cell is a fluidized bed design; the superheater and the convection pass are conventional designs. The two modules will be shipped separately to the construction site for installation. The detailed boiler design is presented in Appendix D.

Boiler. One-inch horizontal tubes are used in the bed with horizontal spacing of 6 in and vertical spacing of 3 in. in staggered positions. A recirculation pump provides the forced circulation of water through the tubes for heat removal. A particle splash screen three tube-rows in depth is incorporated in the design using one-inch tubes on a 3 in x 3 in spacing. Free volumes are provided between the submerged surface and the grid (12 in in depth), and between the top row of the submerged surface and the stop-splash screen (8 in in depth). The freeboard above the stop-splash screen is left empty for gas passage and for easy access to boiler interior. The use of one-inch tubes instead of larger tubes in the bed has the following design advantages: smaller vertical headers can be used; fewer pipes are required to collect the smaller vertical headers into the steam drum, and this in turn means fewer on-site welds; larger heat transfer surface area can be submerged in the bed with a constant tube pitch/tube diameter ratio.

Two-inch tubes on three-inch centers are used for water walls with natural circulation. The same water walls are also used to cool the grid plate.

The high carbon content ash collected in primary cyclones will be recycled back into the carbon burn-up cell running at a higher temperature and excess air.

Carbon Burn-Up Cell (CBC). The presently proposed carbon burn-up cell design was based primarily on turn-down considerations. During turn-down, the bed temperature in the CBC has to be kept constant for efficient carbon burn-up. This becomes a problem in actual operation

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if the CBC is used for superheating the steam. When the boiler is turned down, the total input of carbon fines into the CBC decreases. There are three ways to maintain the bed temperature constant in this turn-down condition: 1) lower the bed depths to decrease the total heat transfer surface in the bed; 2) feed additional fresh coal into the CBC while keeping the bed depth constant; 3) sectionalize the CBC so that sections can be turned off corresponding to the operation of the primary beds. However, methods 1 and 2 raise problems. For example, for a load reduction to 25% of the full load, the superheater metal temperature would be very close to the bed temperature because 25% of the original steam flow would be running through the superheater. Method 1 would require reduction of the bed depth to almost 25% of the designed bed depth -- too shallow to operate efficiently.

The logical solution is to divide the CBC into four separate sections (each corresponding to a primary bed) by installing water tube screens in the bed and by subdividing the air plenum. Turn-down by shutting down a bed would require shutting down the corresponding section in the CBC. In this way, each section would only require a load reduction capability similar to that of each primary bed. This load reduction can be provided by designing the CBC at higher excess air. During load reduction, the excess air is correspondingly reduced to keep the bed temperature in the CBC constant. For 30% load reduction, a reduction of excess air from the design value of 50% to ∿15% is necessary.

Dividing the CBC into two sections is not considered feasible. First, each section would require > 50% load reduction capability, so that the CBC would have to be designed at even higher excess air. That represents an increase in sensible heat loss from flue gas and a decrease in overall efficiency. Second, such a division might complicate the control problem during operation.

Superheater. The superheater is a conventional design and is located in the convection zone right before the convection section. The proposed superheater design occupies approximately 14 feet of the 30-foot convection section, with serpentine superheater tubes arranged for counter-flow of steam to flue gas. Carbon steel tubing of 2-1/2 in o.d. is proposed with 6-1/2-inch perpendicular (to the flue gas flow direction) and 4-in parallel spacing, as shown in the inset of Figure 1-69. With the present design, the control of the superheated steam temperature is expected to be minimal because the flue gas flow rate through the superheater decreases proportionally, as does the heat transfer coefficient, as the load is reduced.

Convection Pass. Two-inch tubes are arranged on 4-1/2 -in spacing in the convection zone to recover heat from the flue gas to a level such that only one heat trap (economizer) is needed further downstream. Natural circulation is employed in this section. No valves are used to parallel flow through various circuits.

Economizer. An extended-surface economizer was chosen for this design on the basis of a study of the cost of heat traps, including tubular and regenerative air heaters, and bare-tube and extended-surface economizers. The extended-surface economizer is a convection unit similar to the bare-tube economizer, with multiple parallel circuits arranged for counter-flow of gas to water. The gas-side heat transfer surface is increased by applying continuous spiral fins 3/4 in high, with two fins per inch of tubing. The tube size considered here is 1-1/2 in arranged in a 4-1/2 in square pitch.

Coal and Limestone Feed Systems

The detailed designs of coal and limestone feed systems are shown in Figure 1-73. These two feed systems are discussed separately below.

Coal Feed System. Altogether 32 coal feed points, eight for each bed, are needed for the four primary beds. This represents about 10 ft^2 bed area per coal feed point, chosen for the present design on the basis of the British data [44]. Four additional feed points are provided for the CBC for feeding the carbon-rich ash collected in primary cyclones. This corresponds to 16 ft^2 bed area per feed point.

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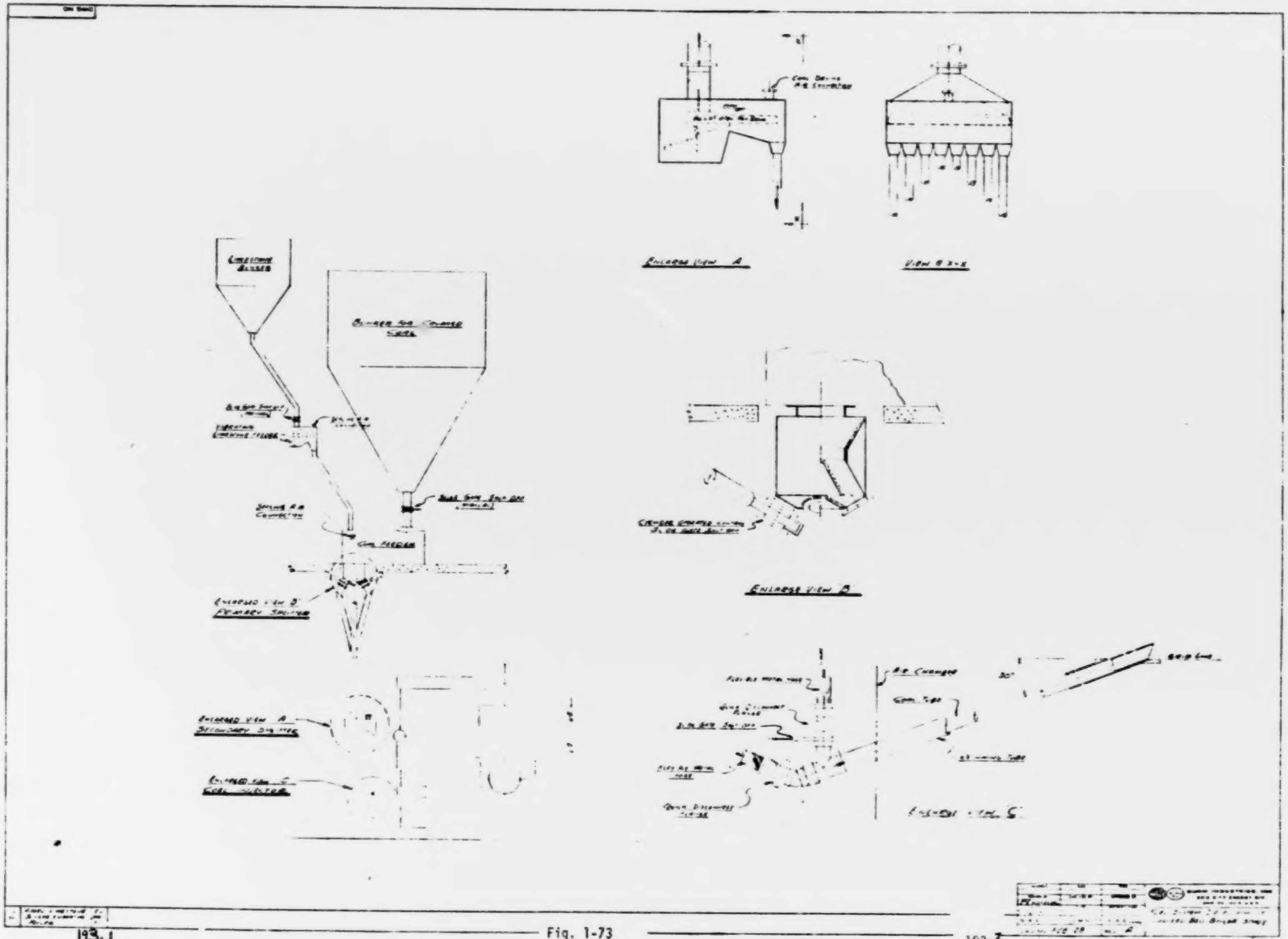
Limestone Bunker
Coal Feeder
Enlarged View B Primary Splitter
Enlarged View A Secondary Splitter
Enlarged View C Coal Injector
Enlarge View A
View @ X-X
Enlarge View B
Air Changer
30°
Enlarge View C

193·1

Fig. 1-73
193·2

193·3

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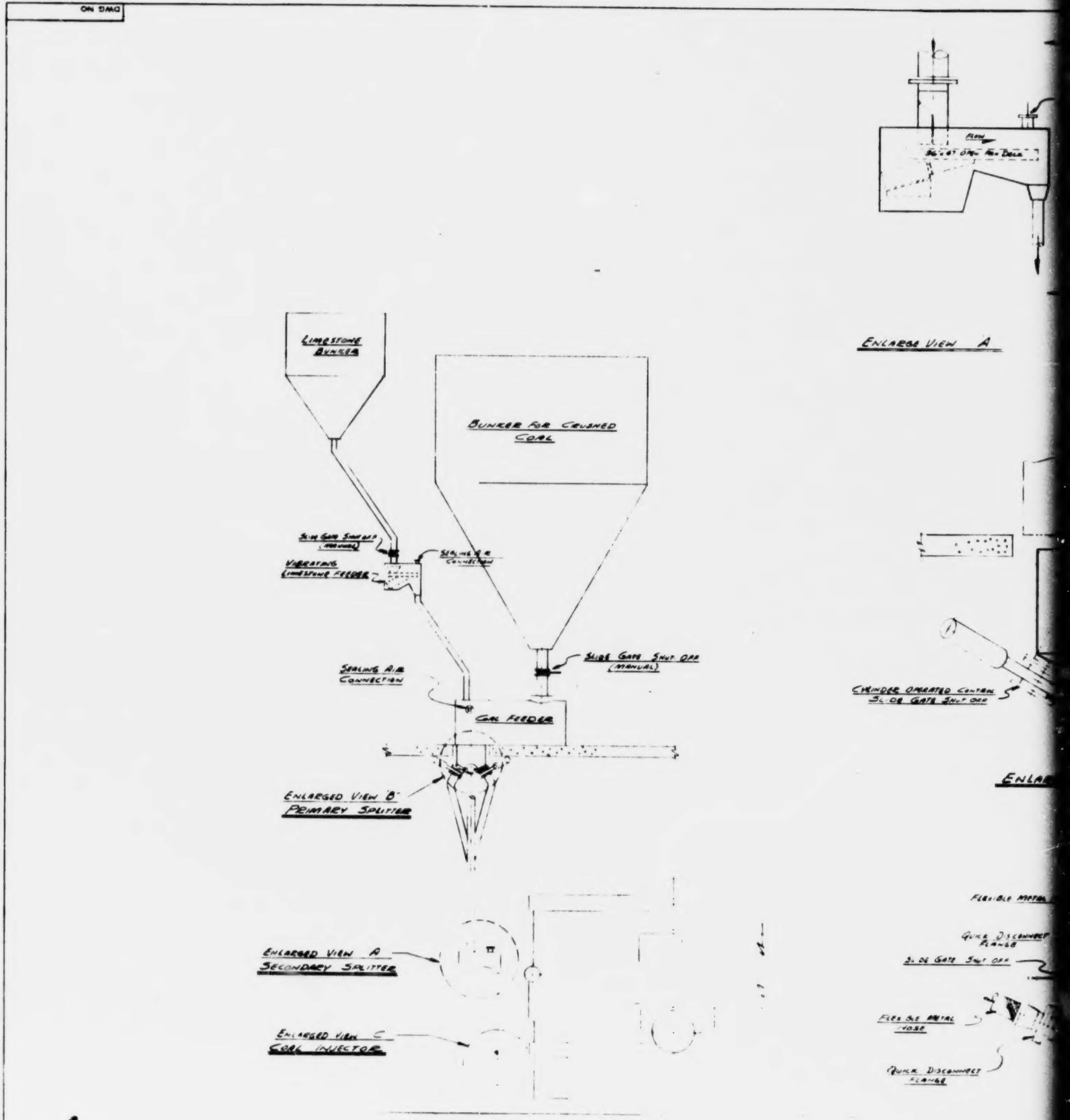
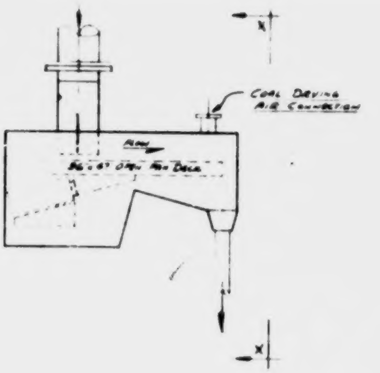

ON DWG
LIMESTONE BUNKER
BUNKER FOR CRUSHED COAL
SLIDE GATE SHUT OFF (MANUAL)
VIBRATING LIMESTONE FEEDER
SEALING AIR CONNECTION
SEALING AIR CONNECTION
SLIDE GATE SHUT OFF (MANUAL)
COAL FEEDER
ENLARGED VIEW "B" PRIMARY SPLITTER
ENLARGED VIEW A SECONDARY SPLITTER
ENLARGED VIEW C COAL INJECTOR
FLOW
ENLARGE VIEW A
CYLINDER OPERATED CONTROL SLIDE GATE SHUT OFF
ENLAR
FLEXIBLE METAL
QUICK DISCONNECT FLANGE
SLIDE GATE SHUT OFF
FLEXIBLE METAL HOSE
QUICK DISCONNECT FLANGE
ADDED LIMESTONE BUNKER VIBRATING FEEDER

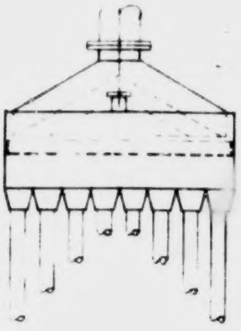
193.1

Fig. 1-73

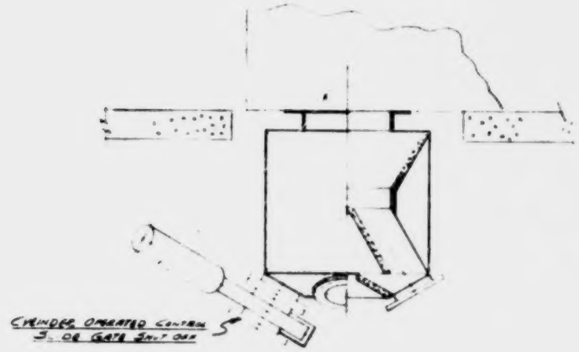
193.2



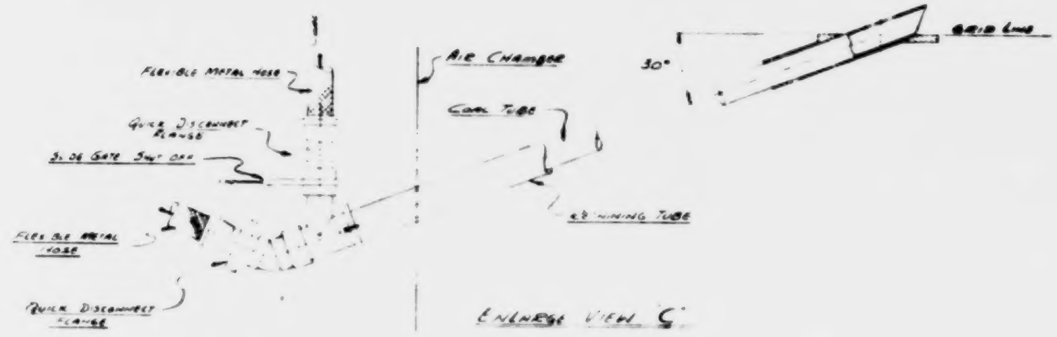
ENLARGE VIEW A



VIEW @ X-X



ENLARGE VIEW "B"



ENLARGE VIEW C

ZURN INDUSTRIES, INC.
ERIE CITY ENERGY DIV.
ERIE, PA. [illegible] U.S.A.

FUEL SYSTEM DEVELOPMENT
FLUIDIZED BED BOILER STUDY

SCALE
DIM NOT TO SCALE
DWG NO FBB-28
REV A

Fig. 1-73
193.2

Coal crushed to 1/4 in (see Figure 1-64) is first transferred to the feeding hopper and then to a moving belt gravimetric feeder with adjustable belt speed for different feed rate control. (See Appendix D for a detailed description of the feeder.) The coal from the gravimetric feeder drops into a four-way splitter to divide the coal into four equal streams, one for each bed. An individual shut-off valve is provided for each stream, but no individual control of the coal feed rate for each stream is possible. Each stream of coal is again split into eight equal portions, one for each coal feed point, by using a vibrating table (detailed in Appendix D). The coal is then fed pneumatically into the bed.

Coal drying is also accomplished in the coal feeding system by using preheated sealing air. No condensation of moisture in the coal feeding lines is expected if the lines are kept at a temperature higher than the dew point of the gas. No SO_2 or particulate will be vented into the atmosphere during this drying operation because of the tight enclosure in the feeding system.

The individual coal feeder is a "straight-in" design forming a 30° angle with the distributor plate to facilitate the maintenance and replacement of the coal feeder even while the boiler is in operation.

Limestone Feed System. Since limestone is mixed with coal at the discharge end of the coal feeder, no additional limestone feed point is provided in the fluid beds. The limestone hopper is a pressure-seal type, and the limestone feed rate is controlled by a variable amplitude vibrating table feeder.

In both the once-through dry solid and the wet scrubbing systems, the limestone feed rate is not regulated according to the coal feed rate but rather is adjusted to keep the SO_x emissions in the stack gas within the regulated limits.

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Particulate Removal System

Primary. Four cyclones of 90% efficiency at 10 µ will be used as primary cyclones. They are refractory lined and internally insulated. Ash collected in the primary cyclones is carbon-rich and is recycled back to the carbon burn-up cell for refiring. At part-load condition the flue gas flow rate is reduced and so is the efficiency of the cyclones. At part-load operation, the overall boiler efficiency can either be sacrificed or be maintained by using high-temperature dampers to bypass some of the cyclones. The latter would be the better solution; however, the high-temperature dampers may not be readily available or may not provide trouble-free service.

Secondary. For the once-through dry solid system, an electrostatic precipitator of 99.5% efficiency (specifications presented in Appendix D) will be used to reduce the flue gas particulate loading. In the wet scrubbing system, the final particulate removal system is incorporated in the wet scrubbers. Zurn Industries' medium-energy wet scrubber system is used here. (The details are shown in Appendix D.)

Ash Handling. Besides the collection of carbon-rich ash in the primary cyclones, the residual ash from the bed sections and from an ash pick-up point underneath the superheater will be pneumatically transported to the ash hopper. In the case of the once-through dry solid system, the ash from the electrostatic precipitator is also discharged to the same ash handling system. (The positive-pressure pneumatic ash-handling system is presented in Figures FBB-30 and FBB-31 in Appendix D.)

Pollution Control System

Four major pollution control systems are under consideration for the industrial boiler application. These are outlined in Table 1-22 with estimations of their respective fresh limestone requirements and total waste solids produced. A complete cost analysis for all four systems was conducted and is presented below. Only two systems, once-through dry solid and wet scrubbing, were selected for detailed design and cost analysis.

TABLE 1-22

COAL-FIRED INDUSTRIAL BOILER POLLUTION CONTROL ALTERNATIVES

BASIS: 4.3 wt % S COAL

SYSTEM	EQUIPMENT FOR EMISSION CONTROL	FRESH LIMESTONE FEED (STOIC. CA/S) (TONS/TON COAL)	WASTE SOLIDS TONS/TON COAL	SO_2 EMISSION PPM (LB/10^6 BTU)	PARTICULATE EMISSION GR/SCF (LB/10^6 BTU)	BOILER SYSTEM PRESSURE DROP (POLLUTION CONTROL SYSTEM, IN H_2O)
Desulfurized char	Electrostatic precipitator	none	~ 0.1			< 0.5
~ 1 wt%S				~ 770	0.01-0.05 (0.02)-(0.08)	
~ 0.3 wt%S				~ 230 (0.52)		
Limestone scrubbing						
• calcine stone in boiler; once-through	Scrubbing systems Water reclamation	~ 1.2/~ 0.18	0.6-0.7	~ 500 (1.13)	0.01-0.05 (0.02)-(0.08)	12-16
• remove SO_2 in bed, regenerate stone, scrub effluent	Scrubbing system Water reclamation Electrostatic precipitator	1-3/~ 0.36	0.6-0.8	~ 500 (1.13)	0.01-0.05 (0-02)-(0.08)	< 0.5
Once-through limestone	Electrostatic precipitator Larger limestone handling facility	~ 6/~ 0.84	~ 0.7	170-350 (0.38-0.80)	0.01-0.05 (0.02)-(0.08)	< 0.5
Limestone regeneration						
• on-site	Electrostatic precipitator Regeneration system	1-2/~ 0.1	~ 0.3	170-350 (0.38-0.50)	0.01-0.05 (0.02)-(0.08)	< 0.5
• off-site central facility	Electrostatic precipitator	1-2/~ 0.1 (assume mixed at central facility)	~ 0.1	170-350 (0.38-0.80)	0.01-0.05 (0.02)-(0.08)	< 0.5

In the once-through dry solid system, six times stoichiometric limestone is fed and calcined in the primary beds. SO_2 removal is expected to be 90% in the bed, which represents about 16% stone utilization. The limestone is used only once through.

About 1.2 times stoichiometric limestone is needed in the wet scrubbing system. The limestone is first calcined in the bed and then mixed with water to form slurry for use in the wet scrubber. Two parallel units of wet scrubbers with two stages each are used for scrubbing the stack gas. The design of the wet scrubber is presented in FBB-29, Appendix D. In addition to the contacting stages of the wet scrubbers, drawing FBB-36 shows the components of the slurry dewatering system, which includes the slurry thickener and vacuum filter. The particulate and SO_2 removal efficiencies for the wet scrubber are expected to be 98% and 90% respectively.

In addition to the wet scrubber design by Zurn Industries described in Appendix D, a budget quote from Norton Company for the packed bed wet scrubber was obtained. Although the packed bed scrubber only costs about 1/3 as much as the wet scrubbing scheme selected, it may suffer more operational problems due to the intrinsic scaling problem of the limestone scrubbing systems and thus is not suitable for the industrial boiler application. (The quote by Norton is included in Appendix D.)

To avoid calcium sulfate precipitates in lines, vessels, and pumps, the flue gas may be scrubbed with a caustic solution. Lime and spent caustic are mixed in a reactor vessel, forming calcium sulfate and regenerating caustic solution. A calcium sulfite, sulfate filter cake is discharged from the process. The Ceilcote Company has prepared a design and budget estimate for such a process, given in Appendix D. It costs about three times more than the selected wet scrubbing system.

Selection of Draft Equipment

The total draft losses for the selected once-through dry solid and wet scrubbing systems are summarized in Table 1-23. A single,

TABLE 1-23

SUMMARY OF DRAFT LOSSES

COMPONENT	AIR RESISTANCE R DRAFT LOSS, IN. WG	
	DRY SOLIDS SYSTEM	WET SCRUBBER SYSTEM
FORCED DRAFT DUCTS	0.2	0.2
FLUIDIZED BED CHAMBER:		
Section dampers	0.5	0.5
Perforated plate	1.0	1.0
Grid	5.0	5.0
Bed	20.0	20.0
MECHANICAL COLLECTOR	5.0	5.0
CARBON BURN-UP CELL (THRU FLOW)	0.5	0.5
SUPERHEATER	2.6	2.6
CONVECTION SECTION	5.6	5.6
ECONOMIZER	2.0	2.0
ECONOMIZER FLUES	0.2	0.2
PRECIPITATOR	0.5	-
PRECIPITATOR FLUES	0.2	-
WET SCRUBBER	-	16.0
SCRUBBER FLUES	-	0.2
TOTAL	43.3	58.8

forced-draft fan or a combination of one forced-draft fan and one induced-draft fan can be used. A cost estimate of both these schemes is presented in Table 1-24. A single forced-draft fan provides the same total pressure drop more cheaply and is also favored in the wet scrubbing system because there is no downstream induced-draft fan to be clogged up by possible deposits of $CaSO_4$.

Overall Design Layout

The proposed overall boiler design arrangements for the once-through dry solid and wet scrubbing systems are shown in Figures 1-74 and 1-75 respectively, with the appropriate dimensions.

TABLE 1-24

DRAFT FANS FOR INDUSTRIAL FLUIDIZED BED COMBUSTION BOILERS[a]

TOTAL ΔP FOR FANS IN H_2O	SINGLE, FORCED-DRAFT		DOUBLE, FORCED-AND INDUCED-DRAFT FANS		
			HP		
	HP	COST, MOTOR + FAN	INDUCED	FORCED	COST, MOTOR + FAN
50.9	800	$33,000	450	500	$46,000
56.9	800	$33,000	450	600	$51,000
59.5	900	$35,000	450	600	$51,000
63.4	1000	$35,000	450	700	$53,000
69.4	1100	$54,000	450	700	$56,000

(a) 250,000 lb/hr industrial boiler
75,000 CFM air flow at 100°F
120,000 CFM top flow at 375°F

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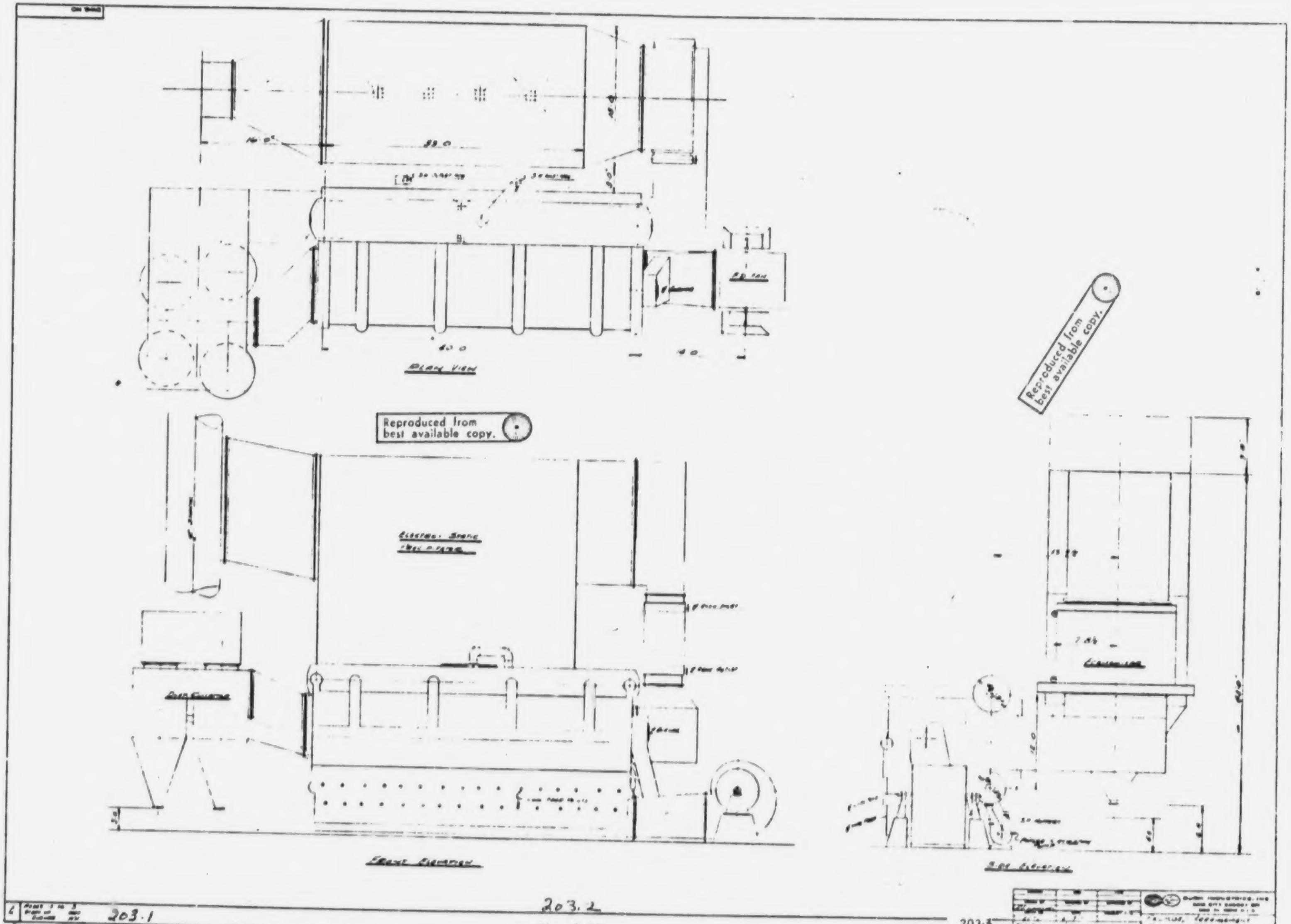

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Fig. 1-74

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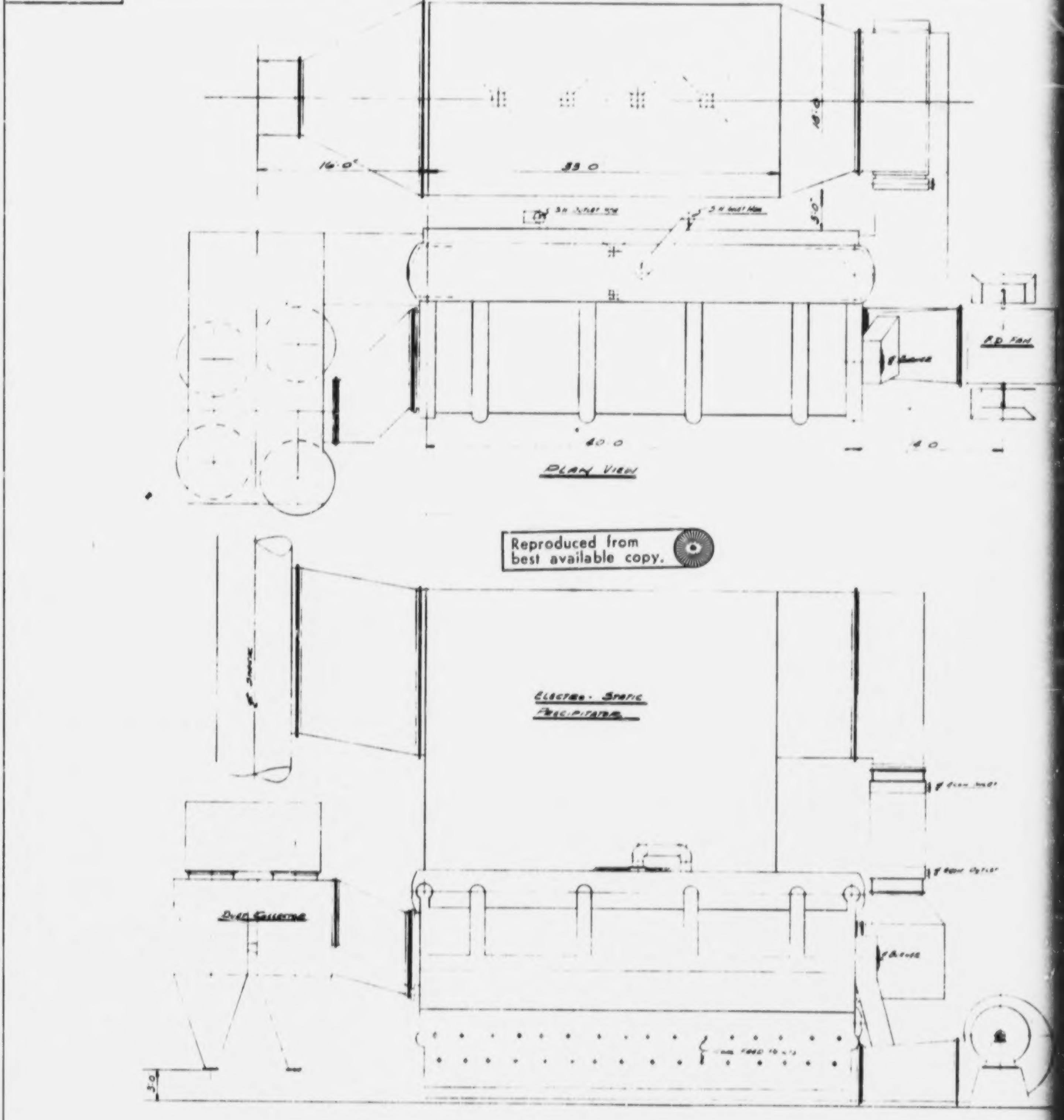

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Fig. 1-74

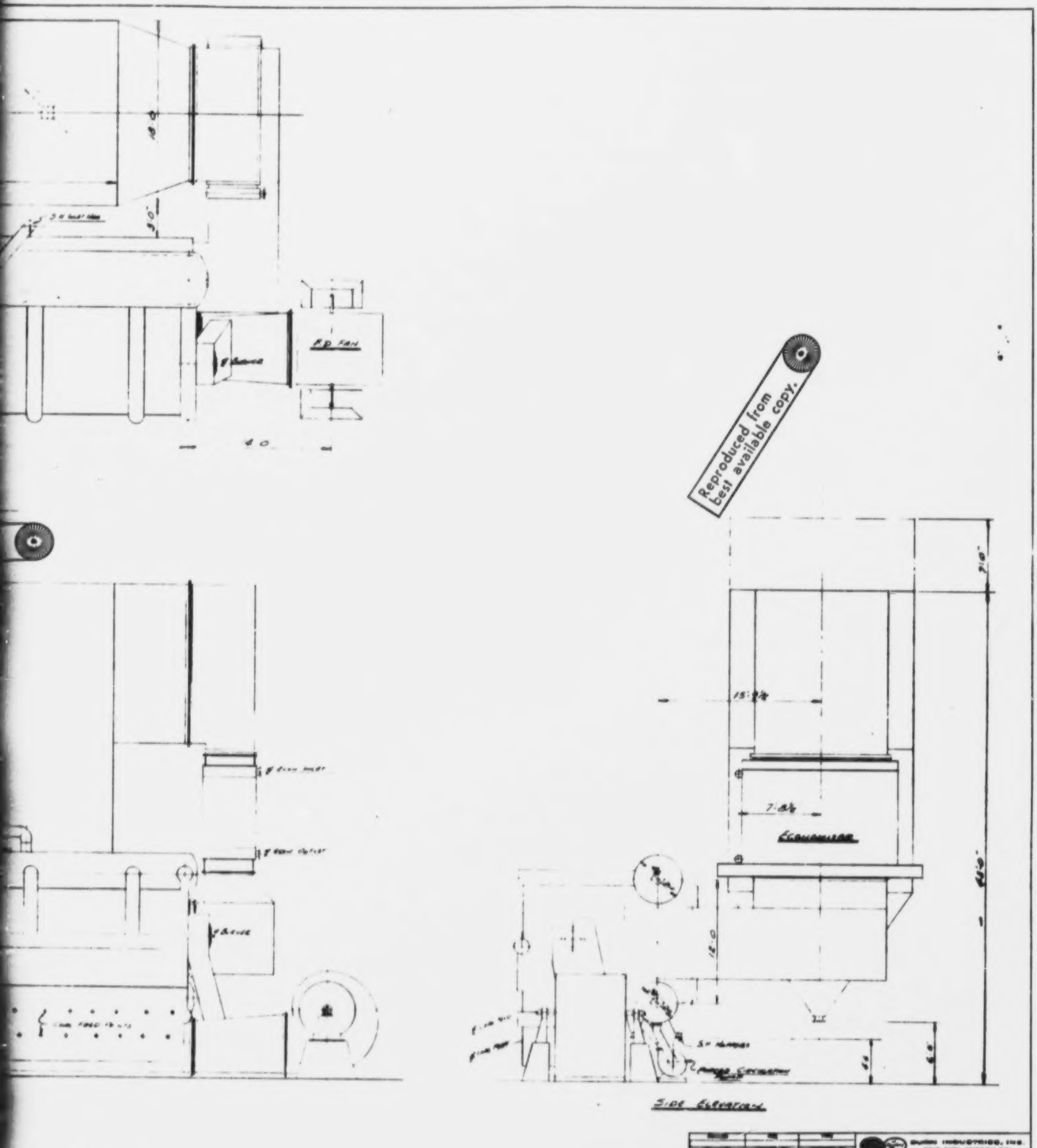

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Fig. 1-74

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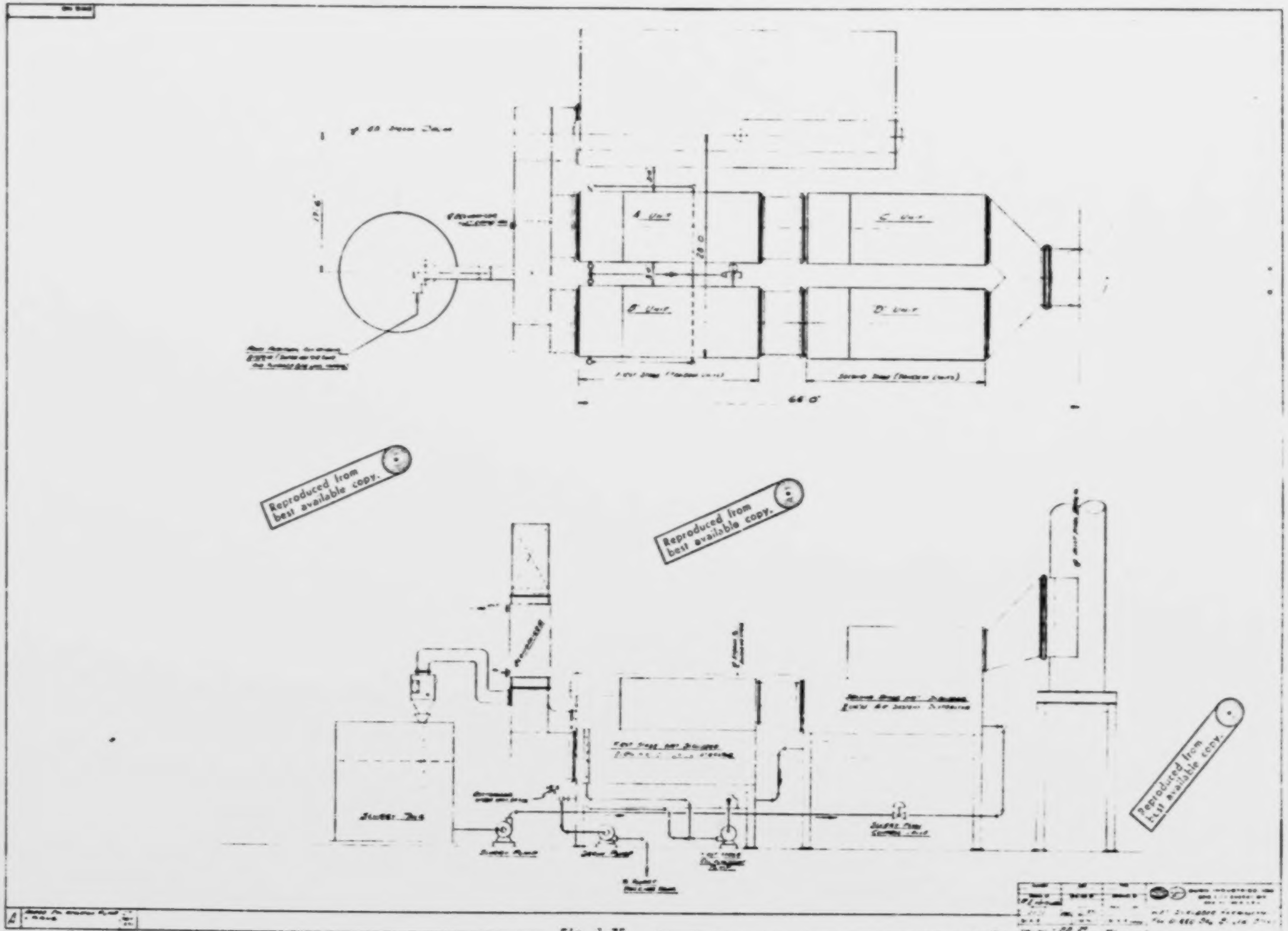

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Fig. 1-75
2 of 2

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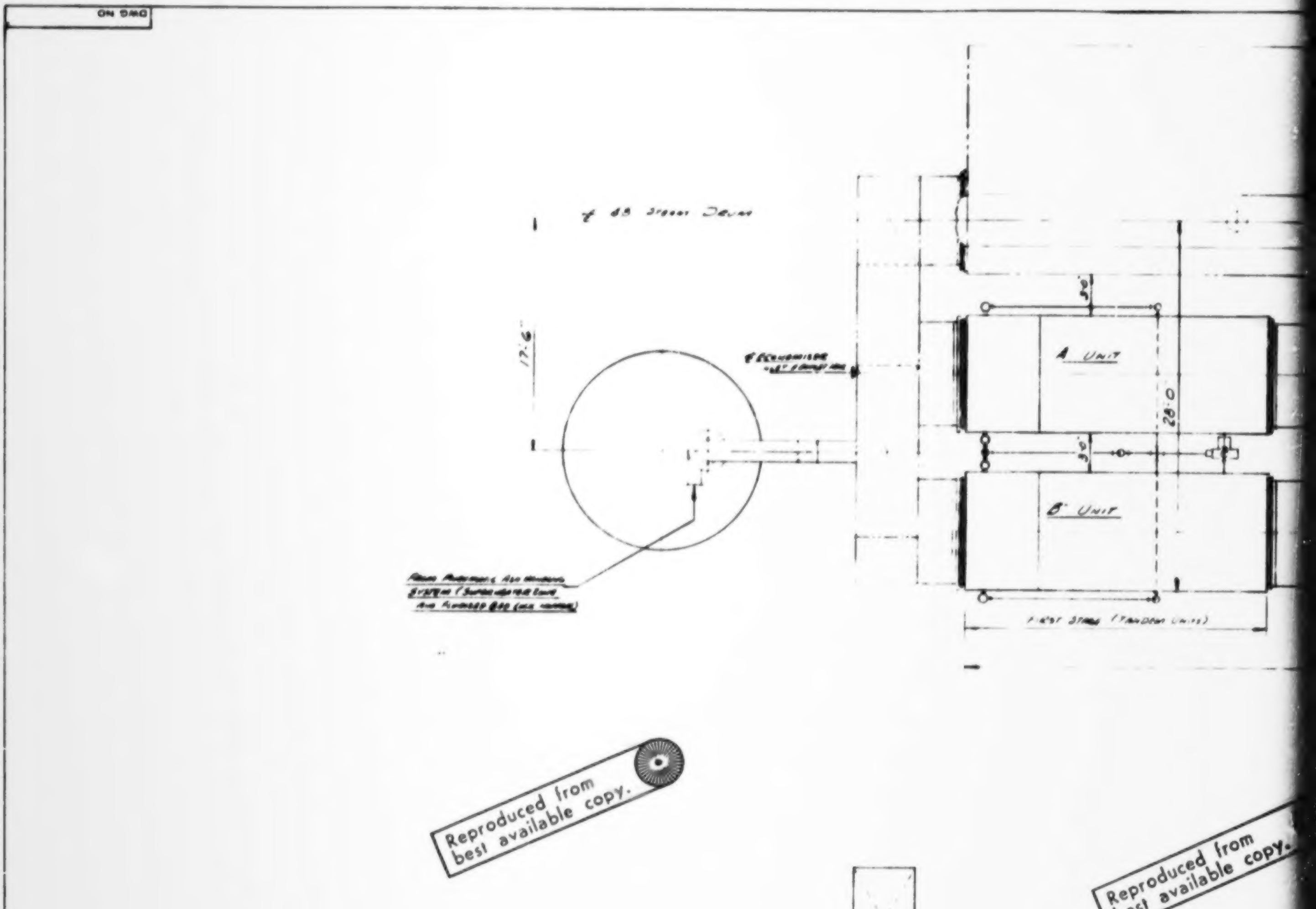
ON DWG
℄ 4B Steam Drum
17'-6"
℄ Economizer
From Pneumatic Ash Handling System (Superheater Zone and Plugged Bed (ash hopper))
A Unit
B Unit
30
28'-0
30
First Stage (Tandem Units)
Economizer
Out
In
℄ Steam & Superheater
Slurry Tank
First Stage Wet Scrubber
Economizer Water Wash Drain
Slurry Pump
Drain Pump
To Slurry Thickener Tank
First Stage Recirculation Pump
A
Added Recirculation Pump & Piping

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Fig. 1-75

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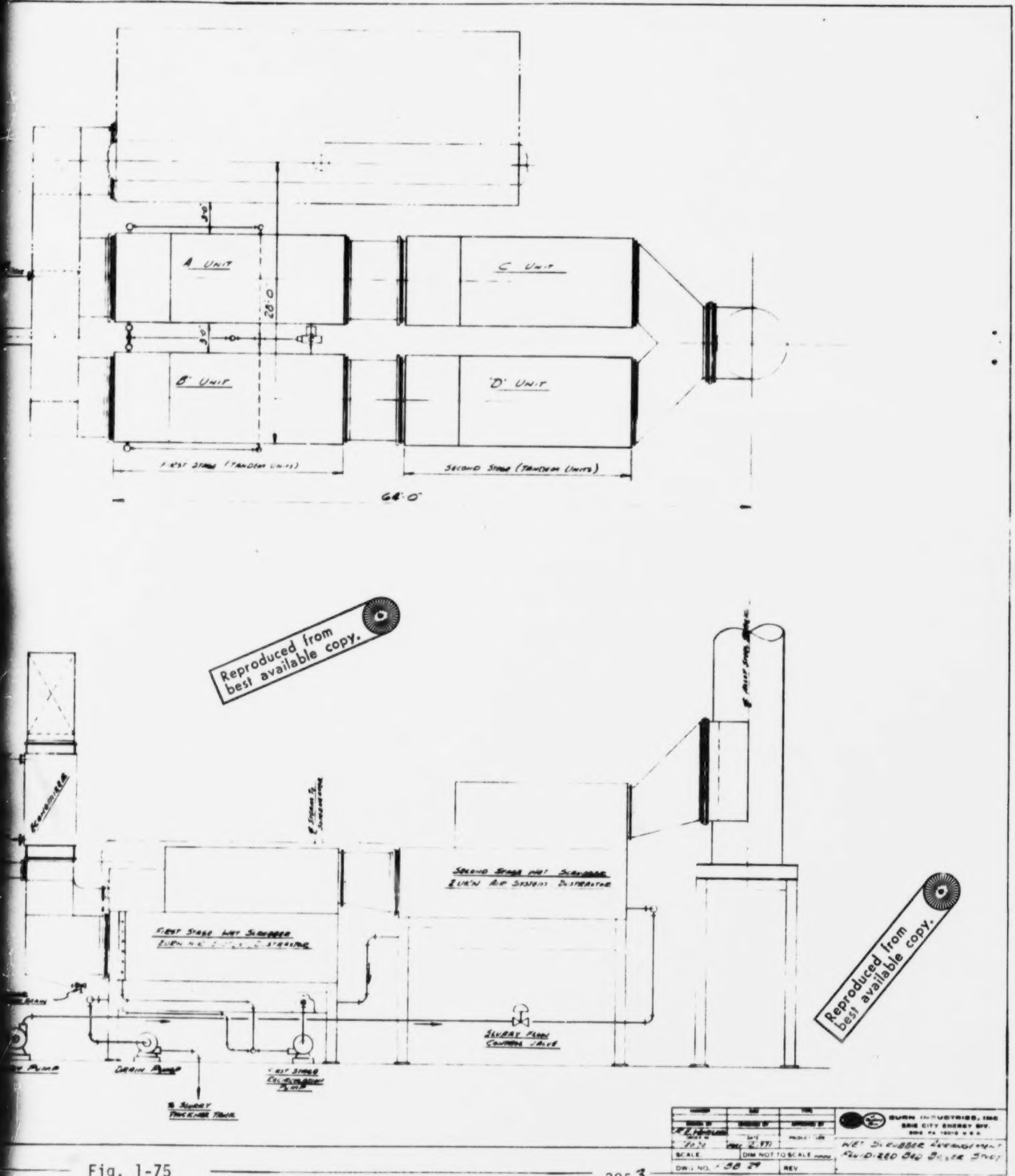

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Fig. 1-75
205-2

Boiler Operation and Performance

A coal-burning fluidized bed boiler would be operated somewhat differently from a conventional chain stoker, and certainly differently from gas or oil-burning boilers. For example, starting up a fluidized bed involves preheating the fluidized bed to a minimum temperature of ~800°F with auxiliary burners before coal can be fed. On the other hand, gas or oil burners for gas- or oil-burning boilers are simply ignited. Providing a turn-down ratio of 4:1 in a fluidized bed boiler necessitates turning on and off sections of fluid beds. Turning off a fluid bed involves only simple procedures; however, turning on a bed once turned off requires preheating the bed to a minimum bed temperature of ~800°F before coal injection, which becomes the limiting factor in load response rate. This requires a separate subcontrol loop. Although a gas- or oil-burning boiler applies the same principles for load control, ignition is much simpler.

Because coal combustion in fluidized beds involves a different combustion mechanism and different operational scheme, the performance characteristics of fluidized bed boilers are necessarily different in some respects from those of conventional boilers. For example, the pressure drop in a fluidized bed is higher because of the relatively high pressure drop through the distributor plate and the fluid bed. In addition to conventional heat losses through the sensible heat of flue gas and moisture in the fuel, the solid carbon loss becomes an important item to consider, due to elutriation.

Although the principles for controlling the fluidized bed boiler are the same as for conventional boilers, the philosophy is different in some respects. The bed temperature is the primary variable for control: too high a temperature will cause ash to fuse and limestone to lose reactivity; too low a temperature will decrease substantially

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the reactivity of limestone and possibly decrease the combustion efficiency of coal in the bed.

The detailed operating procedures, performance characteristics, controls, and instrumentation are discussed below.

Operating Procedures

Start-up

Raising Boiler Pressure to On-Line Conditions. The boiler is started up first by firing an auxiliary gas burner above the bed in the freeboard (as shown in Figure FBB-23 in Appendix D), and raising the boiler pressure to on-line conditions, i.e., 600 psig and 489°F. The bed temperature at this point will be essentially 489°F because of recirculation of saturated water through the immersed horizontal tubes in the bed. The time required to raise the boiler pressure to on-line conditions is estimated to be about 1-1/2 to 2 hours.

Heating Bed Material to Minimum Coal Injection Temperature. Experiments by PER and NCB found a minimum bed temperature of ~800°F to be necessary for self-sustained combustion of coal particles in the fluidized beds. Therefore, the bed must be heated up from the saturation temperature 489°F to at least 800°F by injection of auxiliary gas fuel into the minimum-fluidized bed before coal feeding is started. Only the first of the four beds and sections in the carbon burn-up cell have to be started in this way. Start-up of the other beds can rely on the solid recirculation from the neighboring operating bed to raise the bed temperature. (This problem has been mathematically analyzed in Appendix F.)

Raising Load and Putting Other Beds in Service. Once the bed reaches the minimum coal injection temperature, the coal feed is started and the load is increased. The first section of the carbon burn-up cell can then be fired in the same way, using the elutriated carbon fines from the operating section as fuel. To be started, the next bed can be fluidized at minimum fluidizing conditions, and the bed material is allowed to circulate from the operating bed to raise the bed temperature to minimum coal injection temperature. The second carbon burn-up cell section can also be started by recirculation of solids from the operating section. The third bed and third CBC section can be lit off in the same fashion and so on until the boiler is in full operation. (The dynamic response of the bed temperature during start-up is presented in Appendix F.)

Load Control

Numerous load control techniques for fluidized bed boilers can be employed to achieve the desirable percentage load reduction. However, the preferred turn-down technique is optimal for ease of control, stability and reliability of operation, degree of turn-down desirable, rate of system response required, efficiency, and economy. (A comprehensive analysis of turn-down techniques is presented in Appendix E.)

Load control in fluidized beds involves turning the beds on and off. In turning off the beds, the load response is fast enough to meet 10%/min. load change as analyzed in Appendix F. However, turning on the idled beds to meet the desirable response rate becomes a big challenge. (A comprehensive analysis is included in Appendix F.)

Turn-Down Operation in Single Bed. In atmospheric industrial boilers, a turn-down capability to less than 25% of full load at a load swing rate of at least 5% per minute is desirable. To meet these requirements, a technique which simultaneously decreases fuel and air input and maintains excess air and bed depth constant is recommended.

The recommendation is based on the general analysis presented in Appendix E. Table 1-25 compares the operating conditions of the two most promising modes: 1) turn-down by decreasing fuel and air input and maintaining excess air and bed depth constant; and 2) turn-down by decreasing fuel input and maintaining air input and bed depth constant. The operating conditions before and after turn-down for both modes are compared, assuming the load reduction is to be accomplished by lowering the bed temperature from 1650°F to 1400°F with bed depth constant.

TABLE 1-25

COMPARISON OF EFFECTIVENESS OF TURN-DOWN TECHNIQUES FOR INDUSTRIAL BOILER

Steam Conditions: 750°F & 600 psig

OPERATING CONDITIONS	BEFORE TURN-DOWN	AFTER TURN-DOWN	
		MODE 1	MODE 2
BED TEMPERATURE, °F	1650	1400	1400
EXCESS AIR, %	10	10	40
FUEL REQUIREMENT (ARBITRARY UNIT)	1	0.69	0.79
SUPERFICIAL FLUIDIZING VELOCITY, FT/SEC	15	8	12
LIMESTONE FEED RATE (ARBITRARY UNIT)(a)	1	1.4	1.6
LOAD REDUCTION IN THE BED, %	--	21.5	21.5
GAS TEMPERATURE, °F			
Leaving combustion chambers	1530	1250	1293
Entering convection pass	1530	1250	1293
Leaving convection pass	672	580	626
Leaving economizer	350	315	352
HEAT LOSS THROUGH FLUE GAS (ARBITRARY UNIT)	1	0.96	1.40
OVERALL MAXIMUM TURN-DOWN ACHIEVABLE WITH ALL BEDS OPERATING,%	--	31	21

(a) The reaction rate for limestone-SO_2 reaction at 1400°F is just about half of that at 1650°F. This means either more reactive dolomite has to be used as absorbant or more limestone has to be fed.

This reduces the total heat transferred in the bed by 21.5% after turn-down in both cases, assuming the heat transfer coefficient in the bed is constant and independent of turn-down operations. However, the total heat transferred in the convection pass and economizer is much lower in Mode 1 turn-down than in Mode 2 due to a reduction of almost 30% of the original flue gas mass flow rate, G, which affects the heat transfer coefficient, h, in the convection section by the relationship $h \alpha (G)^{0.61}$. This results in a maximum achievable turn-down in operating beds of 31% for Mode 1 versus 21% for Mode 2. In addition to this advantage, less fuel and a lower limestone feed rate are required, and much less heat is lost through flue gas in Mode 1. The temperature of 1400°F is chosen on the basis of the SO_2 removal capability of limestone and does not represent the minimum operating temperature. For flexibility and convenience in load swing, the bed can be operated at temperatures higher than 1650°F (operation at 1700°F provides ∿7% extension in load) or lower than 1400°F (operation at 1300°F provides an additional ∿8% reduction in load) for short periods of time without seriously affecting efficiency of carbon combustion and SO_2 removal. When wet scrubbing is used for SO_2 removal, the bed temperature is more flexible because it is no longer limited by limestone reactivity in the bed but by ash fusion temperature (∿2000°F) at the upper limit and coal combustion efficiency (∿1200°F) at the lower limit.

Change of bed depth as an alternative means for turn-down was ruled out on the basis of the following considerations:

- The designed bed depth in the atmospheric industrial boiler is already quite shallow, i.e., 2-1/2 ft. A more than 25% decrease in bed depth would increase the carbon carry-over and thus decrease overall efficiency.
- The heat transfer surface exposed by changing the bed depth will still have a high heat transfer coefficient, comparable to that of heat transfer surface in the bed. Hence, the effectiveness of changing the bed depth in turn-down is doubtful.

- The additional cost in providing the rapid transport of the solid in and out of the bed with respective flow control devices does not merit consideration in an atmospheric industrial unit. The stability and reliability of operating a shallow bed with rapid flow of solids in and out of the bed might be poor.

Turn-down Operation in Entire Boiler. The present industrial boiler is sectioned into four equal beds with the flue gases from the individual beds collected before passing through the superheater and a common convection section. The proposed philosophy for load reduction is best illustrated in Figure 1-76: all four beds are simultaneously reduced in load at least 25% before turning off bed number 4 and boosting the other beds to full operation; the remaining three beds are turned down to their lowest load before shutting down the number 3 bed; and so on. The scheme is continued until only the number 1 bed is in operation. The number 1 bed operating at lowest load represents ∿17.5% or ∿10% of total boiler load, depending on the pollution control alternative employed. The operating conditions before and after turn-down for Mode 1 are presented in Table 1-26 for beds operating in the temperature range 1400°F to 1650°F. It is significant that the heat loss through flue gas is actually decreasing -- or the boiler efficiency is increasing -- during turn-down; this is because the size of the common convection section is designed for full-load operation. This same amount of heat transfer area is also available during turn-down conditions with a lower flue gas flow rate, resulting in lower stack gas temperature, as shown in Table 1-26. The fact that the efficiency of the boiler is actually increasing during turn-down provides additional flexibility for load reduction. Now Mode 2, which achieves turn-down by increasing excess air, can also be incorporated to give more flexibility in turn-down without suffering loss of boiler efficiency. In the actual boiler operation, the heat loss through the flue gas might be larger than that shown in Table 1-26 because of by-passing of some flue gas into the cold beds which are not in operation. This is especially true when three of

Curve 643916-A

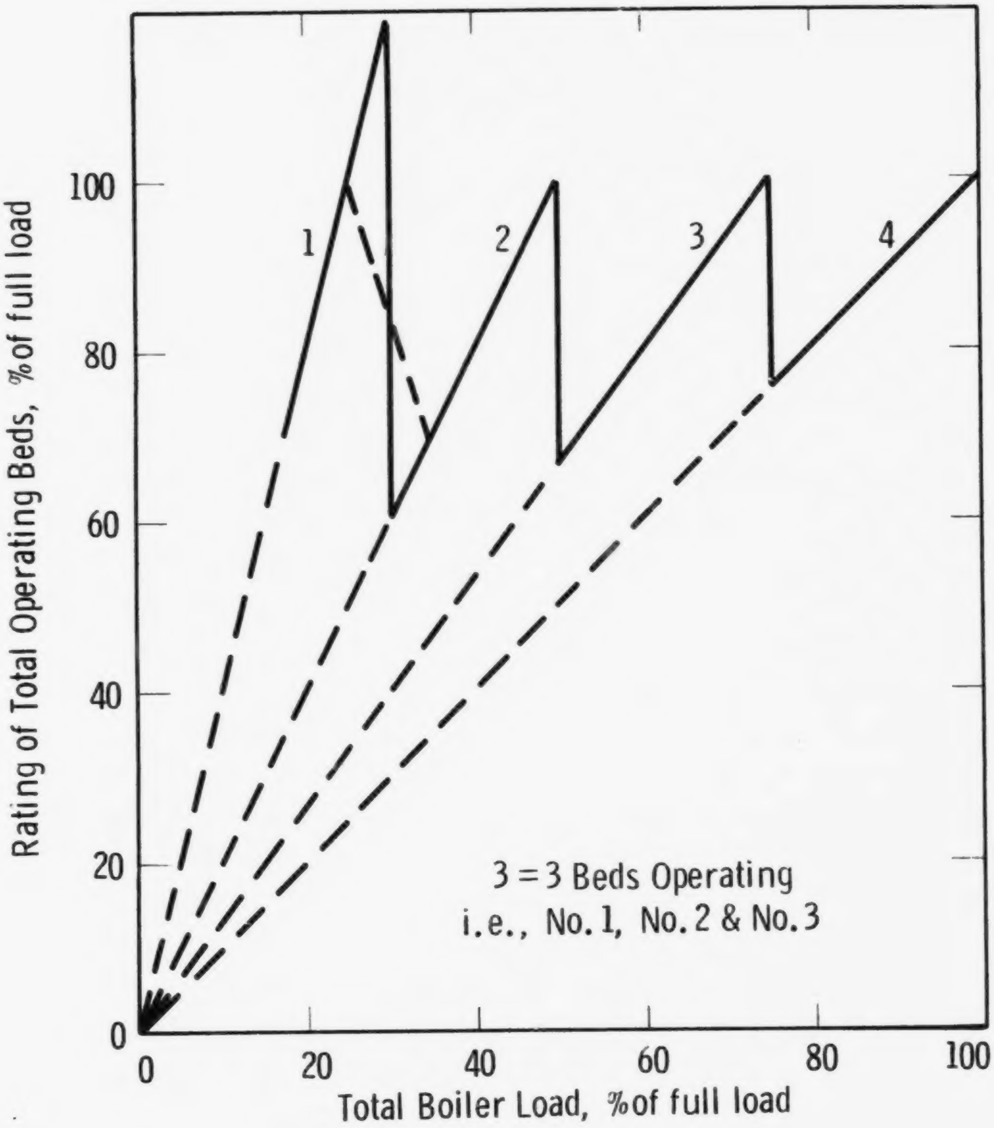

Fig. 1-76–Turn-down scheme of four bed industrial boiler

TABLE 1-26

OPERATING CONDITIONS DURING TURN-DOWN OPERATION FOR INDUSTRIAL BOILER

OPERATING CONDITIONS	4 BEDS OPERATING		3 BEDS OPERATING		2 BEDS OPERATING		1 BED OPERATING	
	Full Load	Part Load	Full Load	Part Load	Full Load	Part Load	Full Load	Part Load
BED TEMPERATURE, °F	1650	1400	1650	1400	1650	1400	1650	1400
EXCESS AIR, %	10	10	10	10	10	10	10	10
FUEL REQUIREMENT (ARBITRARY UNIT)	1	0.69	0.75	0.52	0.50	0.35	0.25	0.17
SUPERFICIAL FLUIDIZING VEL. FT/SEC	15	8	15	8	15	8	15	8
LIMESTONE FEED RATE (ARBITRARY UNIT)	1	1.4	0.75	1.05	0.50	0.70	0.25	0.35
VELOCITY IN CONVECTION PASS (ARBITRARY UNIT)	1	0.59	0.74	0.45	0.50	0.31	0.25	0.15
GAS TEMPERATURE, °F								
Leaving combustion chamber	1530	1250	1516	1281	1516	1281	1521	1281
Entering convection pass	1530	1250	1516	1281	1516	1281	1521	1281
Leaving convection pass	672	580	634	568	592	543	540	513
Leaving economizer	350	315	335	305	310	290	280	270
HEAT LOSS THROUGH FLUE GAS (ARBITRARY UNIT)	1	0.96	0.96	0.85	0.87	0.79	0.76	0.72
PERCENTAGE OF TOTAL LOAD, %	100	69.0	75.5	51.8	50.8	34.8	25.7	17.7

the four beds are shut down; however, the extent of the heat loss is hard to estimate. On the other hand, the flue gas velocity in the convection zone is reduced up to 85% of the original velocity during turn-down and may cause the deposition of ash in the convection section. Erie City's experience indicates this will not be an operational problem.

Turn-Down Operation in Carbon Burn-Up Cell (CBC). The bed temperature in the carbon burn-up cell has to be kept constant during turn-down for efficient carbon burn-up. The proposed design divides the CBC into four separate sections, each corresponding to a primary bed, by installing water tube screens in the bed and by subdividing the air plenum. Turn-down by shutting down a bed would require shutting down the corresponding section in the CBC. Each section in the CBC thus has to provide only the same load reduction capability as each primary bed. The load reduction in the CBC can be accomplished by designing the full-load operation conditions at 1900°F and high excess air, say 50%. When the CBC is operated at reduced load, the percentage excess air can be decreased to decrease the heat loss to the flue gas and keep constant the bed temperature of the CBC. For a 30% load reduction, the operating conditions of the CBC after turn-down are 1900°F and ∿15% excess air. To operate the CBC in this way, individual on-off control of the spent-coal feed to each section of the CBC is provided.

Shut-down

Single Bed Shut-Down. When removing one bed from service, the coal feed is shut off first while fluidizing air is maintained to burn off the residual coal in the bed and to prevent deposit of coal in the pneumatic transport lines. After the residual coal is burned out, the bed temperature starts to drop, and the fluidizing air can be shut off. However, the recirculation water-flow through the immersed horizontal tubes will be maintained throughout the operation. Thus, the heat removal from the defluidized bed after shut-down depends entirely on conduction of bed material to the water tubes. (The change of bed temperature and the residual steam production after shut-down is analyzed mathematically in Appendix F.)

Entire Boiler Shut-Down. To shut down the entire boiler, one fuel-burning bed section is shut down at a time. The start-up burner should be ignited prior to shutting down the last bed section to maintain the steam generator well above the flue gas condensation temperature. The cooling rate is limited so that there are no undue thermal stresses. A cooling rate of 100°F per hour, the same as the heating rate, should be used.

The recirculating pump is kept running until the steam generator is cooled to below 200°F and is ready for draining. As long as the recirculating pump is operating, the water level in the steam drum must be maintained.

Emergency Operation. Foreseeable emergency situations and their remedial procedures are outlined below:

1. Loss of combustion air flow
   - Trip all fuel feed to the steam generator
   - Continue operation of sealing air fan and ash system blowers
   - Continue operation of recirculating pump and maintain water level in steam drum
   - Open superheater vent as soon as boiler pressure drops below the line pressure.
2. Loss of recirculation pump
   - Trip air and fuel feed
   - Continue operation of sealing air fan, ash-handling equipment, and gas clean-up equipment
   - Maintain steam-drum water level
   - Open superheater vent when boiler pressure drops below the line pressure.

3. Loss of steam-drum water level
   - Stop fuel feed and cool steam generator as rapidly as possible
   - Continue operation of forced-draft fan
   - Observe operation of recirculation pump to prevent cavitation
   - Drum level should be restored by feeding water at a low rate (5% to 10% of the maximum steam flow).
4. Sudden burn-out of bed tubes
   - Shut down steam generator as soon as possible.
5. Loss of bed fluidization
   - Trip fuel and then air feed
   - Continue operation of recirculation pump.
6. Loss of fuel feed is not in itself an emergency condition as long as the forced-draft fan and other equipment continue to operate. However, care should be taken to prevent damage to the superheater due to over-firing on restarting.

Performance Characteristics

Overall Boiler Efficiency. The comparison of total boiler efficiency between once-through dry solid and wet scrubbing systems is presented in Table 1-27. The wet scrubbing system has higher boiler efficiency because the stack gas temperature is 132°F, compared to 350°F in the dry solid system. An additional efficiency loss of about 5% in the wet scrubbing system is expected if the flue gas is to be heated up from 132°F to 350°F.

Draft Losses. A summary of draft losses for the once-through dry solid system and the wet scrubbing system is presented in Table 1-28, and a comparison of corresponding power requirements for both systems is estimated in Table 1-29.

TABLE 1-27

COMPARISON OF BOILER PERFORMANCE

	ONCE-THROUGH DRY SOLID(%)	WET SCRUBBING(%)
Loss due to calcination and SO_2 absorption	2.74	0.56
Loss due to water and hydrogen in fuel	3.80	3.83
Loss due to unburned carbon	1.93	2.05
Loss unaccounted for and due to radiation	1.50	1.50
Loss due to evaporation in wet scrubber	--	4.84
Loss due to flue gas sensible heat	6.61 (Flue gas temp.=350°F)	1.30 (Flue gas temp.=132°F)
Total loss	16.58	14.08
Overall boiler efficiency	83.42	85.92

TABLE 1-28

SUMMARY OF DRAFT LOSSES, IN. H_2O (GAUGE)

COMPONENT	DRY SOLIDS SYSTEM	WET SCRUBBER SYSTEM
FORCED DRAFT DUCTS	0.2	0.2
FLUIDIZED BED CHAMBER:		
Section dampers	0.5	0.5
Perforated plate	1.0	1.0
Grid	5.0	5.0
Bed	20.0	20.0
MECHANICAL COLLECTOR	5.0	5.0
CARBON BURN-UP CELL (THRU FLOW)	0.5	0.5
SUPERHEATER	2.6	2.6
CONVECTION SECTION	5.6	5.6
ECONOMIZER	2.0	2.0
ECONOMIZER FLUES	0.2	0.2
PRECIPITATOR	0.5	-
PRECIPITATOR FLUES	0.2	-
WET SCRUBBER	-	16.0
SCRUBBER FLUES	-	0.2
TOTAL	43.3	58.8

TABLE 1-29

COMPARISON OF POWER REQUIREMENTS(a)

STEAM GENERATOR	COAL-FIRED FLUIDIZED BED COMBUSTION	
	DRY SOLIDS SYSTEM	WET SCRUBBER
FORCED-DRAFT FAN, HP	700	950
RECIRCULATING PUMP, HP	200	200
COAL CRUSHER, HP	60	60
SLURRY PUMPS, HP	-	150
TOTAL POWER REQUIREMENTS, HP	960	1360
POWER REQUIREMENTS % OF OUTPUT (b)	0.86	1.21

(a) Only major operating horsepower requirements are shown.

(b) Where a horsepower hour is 2544 Btu and output is 289 million Btu per hour.

Air Pollution Control Capability. The control capability of the industrial fluidized bed boiler for NO_x, SO_2, and particulates is summarized in Table 1-30 from the existing experimental data discussed previously. NO_x emissions from the industrial boiler are expected to be in the range 250 ppm (0.40 $lbNO_2/10^6BTU$) to 500 ppm (0.81 $lbNO_2/BTU$), compared to ∿550 ppm from a conventional spreader stoker. SO_2 removal depends primarily on the amount of limestone fed and is expected to be >90% for the once-through dry solid system. The wet scrubbing system is expected to have an SO_2 removal efficiency of ∿90%. For a coal containing 4.3 wt% sulfur, the SO_2 emission is in the range of 350 ppm to 500 ppm (0.80-1.13 $lb/10^6BTU$). For the once-through dry solid system, the particulate loading in flue gas is expected to be less than 0.02 gr/SCF. The particulate removal efficiency of the present wet scrubber design is ∿98%.

Controls and Instrumentation

Overall boiler control has three subdivisions: combustion control, feedwater flow control, and steam temperature control. The basic control principles for the fluidized bed boiler are similar to those for any subcritical recirculating boiler. The detailed control loops are presented in Appendix D, along with lists of control equipment; the functional aspects of the control logic are outlined here.

Combustion Control System. Steam drum pressure is the load-sensitive control element and is used to initiate the fuel feed. Combustion air is fed to a common plenum by the forced-draft fan, with the plenum air pressure kept constant by the inlet vanes positioned by the signal from the fuel:air ratio setter. Individual dampers control the combustion air flow to each individual primary bed and each individual section of the carbon burn-up cell. Flow to each section is measured and totalized. An oxygen analyzer for flue gas will be provided to record the excess air, but it will not be in the control loop.

TABLE 1-30

AIR POLLUTION CONTROL CAPABILITY
OF INDUSTRIAL FLUIDIZED BED BOILER

	DRY SOLID	WET SCRUBBING
NO_x emission		
ppm	250-500	250-500
$lb/10^6 BTU$	0.40-0.81	0.40-0.81
SO_2 emission		
ppm	350 (90% removal)	350-500
$lb/10^6 BTU$	0.80	0.80-1.13
Particulate emission		
gr/SCF	0.01-0.05	0.01-0.05
$lb/10^6 BTU$	0.02-0.08	0.02-0.08

Temperature in each bed is measured and used as a limit in starting up and shutting down an individual bed. Bed temperature in the carbon burn-up cell is always kept constant and is used as a sensitive control element for adjusting the excess air in the CBC.

For bed start-up a sub-loop, including a transfer switch and time-delay volume chamber, is used to control the rate of air flow introduction and bed fluidization.

Feedwater Flow Control System. The feedwater flow control, or steam drum level control, has three elements. Feedwater and steam flow signals are compared. If a difference occurs, a signal is sent to a rate response system and to the feedwater flow control. The steam drum level is compared to a set point. If a difference exists, a signal is sent through a summer to the feedwater control system. The signals from the flow and from the level are compared and sent to a proportional reset controller, to the feedwater hand-auto station, and then on to the flow control valve positioner.

Steam Temperature Control System. The steam temperature control system will be a two-element system. The outlet steam temperature signal is compared to a set point. The resulting error signal is sent through a proportional plus reset controller and summed with the steam flow signal. The combined signal is passed through the steam temperature attemperator hand-auto station to the spray flow control valve.

Boiler Cost

The capital costs of the two designs selected (once-through dry solid and wet scrubbing systems) were evaluated by using the regular boiler pricing procedures employed by Erie City. The resulting boiler capital cost was then compared with the cost of the conventional coal-fired spreader stoker and gas- or oil-fired industrial boiler with approximately the same capacity and steam conditions.

The capital costs of other designs with different pollution control schemes (shown in Table 1-22) were also derived from these two basic designs, as were the capital costs of gas- and oil-fired fluidized bed boilers.

The total costs of all the alternatives, including labor, power, solid disposal, fuel, and limestone costs were then compared to determine the better systems for industrial application.

Capital Cost of Once-Through Dry Solid and Wet Scrubbing Industrial Fluidized Bed Boilers

The fluidized bed steam generator design described previously was subjected to the standard boiler pricing procedures employed by Erie City. The auxiliary equipment costs for once-through dry solid and wet scrubbing systems were procured. The installation cost of the individual equipment was also estimated. These were used to determine the capital costs of the once-through dry solid system, shown in Table 1-31 and of the wet scrubbing system, shown in Table 1-32. The steam generator cost represents about 45% to 55% of the overall boiler system cost. Aside from the steam generator, the ash-handling system is the major cost item in the dry solid system (17%); and the wet scrubber system is the major cost item in the wet scrubbing system (22%). Surprisingly enough, the fuel feeding system represents only ~6% of the total boiler cost.

TABLE 1-31

INSTALLED COST OF ONCE-THROUGH DRY SOLID SYSTEM

(Shipped As Two Shop-Assembled Modules)

	APPROXIMATE EQUIPMENT COST $	APPROXIMATE INSTALLATION COST $	TOTAL $	% OF TOTAL COST
Crusher	18,000	not included	18,000	1.18
Fuel system (not including bunkers, and system upstream of bunkers).	74,000	25,000	99,000	6.50
Forced draft fan	35,000	5,000	40,000	2.62
Fuel system sealing fan & heater system	12,000	10,000	22,000	1.44
Steam generator system, recirculating pump & economizer	753,000	95,000	848,000	55.64
Ash-handling system	170,000	85,000	255,000	16.73
Mechanical dust collector	30,000	15,000	45,000	2.95
Controls	61,000	30,000	91,000	5.98
Electrostatic precipitator	76,000	30,000	106,000	6.96
TOTAL	1,229,000	295,000	1,524,000	100.00

TABLE I-32

INSTALLED COST OF WET SCRUBBER SYSTEM

(Shipped As Two Shop Assembled Modules)

COMPONENT	APPROXIMATE EQUIPMENT COST $	APPROXIMATE INSTALLATION COST $	TOTAL $	% OF TOTAL COST
Crusher	18,000	Not included	18,000	0.96
Fuel system (not including bunkers, and system upstream of bunkers)	74,000	25,000	99,000	5.35
Forced draft fan	54,000	5,000	59,000	3.19
Fuel system sealing Fan & heater system	12,000	10,000	22,000	1.19
Steam generator system Recirculating pump & economizer	753,000	95,000	848,000	45.79
Ash system for slurry system for carbon burn-up	100,000	50,000	150,000	8.10
Mechanical dust collector	30,000	15,000	45,000	2.43
Wet scrubber system	300,000	100,000	400,000	21.60
Controls	61,000	30,000	91,000	4.91
Dewatering	80,000	40,000	120,000	6.48
TOTAL	1,482,000	370,000	1,852,000	100.00

Comparison of Capital Cost Between Fluidized Bed Boilers and Conventional Boilers

The installed cost of the two basic fluidized bed boilers was compared to that of a field-erected conventional coal-fired spreader stoker and field-erected gas- or oil-fired boilers with equivalent capacity, as shown in Table 1-33. The bases for comparison are presented in the description following that table. The once-through dry solid fluidized bed boiler is cheaper, with an installed cost of $6.10/lb/hr as compared to $7.40/lb/hr for the wet scrubbing system and $7.57/lb/hr for a conventional coal-fired spreader stoker. However, the field-erected gas- or oil-fired boiler cost only $2.44/lb/hr. If the gas- or oil-fired steam generator is shop-assembled, the installed cost can be reduced an additional 16%.

The capital costs for other pollution control alternatives and the gas/oil-fired fluidized bed boiler were also derived from these basic cases and compared in Table 1-34. Some schemes, such as once-through limestone injection in a conventional spreader stoker and in a field-erected oil-fired boiler burning high-sulfur oil, are clearly not optimum schemes because of low sulfur removal capability (∿40%), but they are included in the table for the sake of completeness.

Comparison of Total Cost Between Fluidized Bed Boilers and Conventional Boilers

The total costs of all the air pollution control alternatives (Table 1-22) (including labor, power, solid disposal, fuel and limestone cost) are compared in Table 1-35 to determine the better systems for industrial application and to evaluate the limitations of each individual system. The total cost was obtained from the assumptions that the capital charge is 16.7% per year with a load factor of 70%; that the coal cost is $7.50/ton (an equivalent of 30¢/10^6 BTU), that sorbent cost is $3.00/ton; and that waste disposal cost is $5.00/ton.

TABLE 1-33

COMPARISON OF INSTALLED COSTS(a) FOR EQUIVALENT-CAPACITY, FIELD-ERECTED, STEAM GENERATORS FIRING DIFFERENT FUELS

STEAM GENERATOR (See detailed description)	A	B	C	D
Fuel & Firing Technique	Gas-or oil-firing	Coal-fired, spreader stoker	Coal-fired, fluidized bed combustion	
SO_2 Emission Control	Low-sulfur fuel	Not furnished	Dry solids system	Wet scrubber system
Installed Cost for Boiler System, $	610,000(b)	1,375,000	1,524,000	1,852,000
Installed Cost for SO_2 Control System, Add, $	- - -	520,000	- - -	- - -
Installed Cost Total for Equivalent Emission to Atmosphere, $	610,000	1,895,000	1,524,000	1,852,000
Cost per Unit Steam Generation, $/#/hr. for Equivalent Stack Emission	2.44	7.57	6.10	7.40

(a) Costs are based on experience with similar steam generating units escalated to current prices. Estimated factory and installation prices are used for FBC Steam Generators.

(b) If the steam generator is shop assembled, the installed cost will be reduced 16%.

DESCRIPTION OF THE FOUR DIFFERENT STEAM GENERATORS USED IN CAPITAL COST AND OPERATING COST COMPARISONS

Steam Generator "A"

Erie City Energy Division, Field-Erected, "Keystone" Steam Generator

OPERATING CONDITIONS:
- Steam flow -- 250,000#/hr.
- Steam pressure -- 600 psig
- Steam temperature -- 750°F
- Feedwater temperature -- 250°F
- Stack temperature -- 300°F
- Boiler efficiency -- 82.2%
- System resistance -- 11.8 in wg.
- (Pressurized operation)

FUEL:
- Natural gas -- #2 oil

TERMINALS ARE:
- Inlets to: economizer, fuel train and stack
- Outlets from: superheater

EQUIPMENT FURNISHED:
- Burners -- ECED circular burners
- Boiler -- bottom supported - no preassembly
- Superheater
- Economizer -- extended surface
- Flues, ducts and dampers (not including stack)
- Supporting steel
- Soot blowers
- Combustion and feedwater control systems
- Refractory and insulation
- Piping for boiler fuel system, connection between economizer and boiler and trim piping

SERVICES FURNISHED:
- Erection supervision
- Erection labor
- Start-up service

Steam Generator "B"

Erie City Energy Division, Field-Erected, "Cross Drum" Steam Generator

OPERATING CONDITIONS:
- Steam flow -- 250,000 #/hr
- Steam pressure -- 600 psig
- Steam temperature -- 750°F
- Feedwater temperature -- 250°F
- Stack temperature -- 400°F
- Boiler efficiency -- 85%
- System air resistance draft loss -- 12.4 in wg.
- (Balanced draft)

FUEL:
- Bituminous coal sized to no more than 30% through 4 mesh screen

TERMINALS ARE:
- Inlets to: economizer, stoker feeders and stack
- Outlets from: ash-handling system and superheater

EQUIPMENT FURNISHED:
- Forced-draft and induced-draft fans and drives
- Stoker -- Detroit roto grate -- spreader stoker
- Boiler -- no preassembly -- top supported
- Superheater
- Economizer -- bare tube
- Boiler columns and top grid steel
- Soot blowers
- Combustion and feedwater controls
- Refractory and insulation
- Mechanical collector
- Electro-static precipitator
- Flues, ducts and dampers (not including stack)
- Ash-handling system
- Piping for connection between economizer and boiler and trim piping

SERVICES FURNISHED:
- Erection supervision
- Erection labor
- Start-up service

STEAM GENERATOR DESCRIPTIONS (Continued)

Steam Generator "C"

Erie City Energy Division, Modular, Fluidized Bed Combustion, Steam Generator Using Dry Solids SO_2 Control

OPERATING CONDITIONS:
Steam flow -- 250,000#/hr.
Steam pressure -- 600 psig
Steam temperature -- 750°F
Feedwater temperature -- 250°F
Stack temperature -- 350°F
Boiler efficiency -- 85.4%
System resistance -- 43.4 in wg.
(Pressurized operation)

FUEL:
Bituminous coal -- 100% through 4 mesh

TERMINALS ARE:
Inlets to: economizer, coal feeder, limestone feeder and stack
Outlets from: ash silo, superheater

EQUIPMENT FURNISHED:
Forced-draft fan and drive
Coal feeding system, including rate feeder, splitter and injectors
Limestone feeder
Boiler -- bottom supported pre-assembled into two modules
Superheater
Economizer -- extended surface
Supports for mechanical collector economizer and precipitator
Combustion and feedwater controls
Refractory and insulation
Mechanical collector
Electro-static precipitator
Flue, ducts and dampers (not including stack)
Ash-handling system and silo
Piping for boiler fuel system, between economizer and boiler and trim piping

SERVICES FURNISHED:
Erection supervision
Erection labor

Steam Generator "D"

Erie City Energy Division Modular, Fluidized Bed Combustion Steam Generator Using A Wet Scrubber - Particulate and SO_2 Removal System

OPERATING CONDITIONS:
Steam flow -- 250,000#/hr.
Steam pressure -- 600 psig
Steam temperature -- 750°F
Feedwater temperature -- 250°F
Stack temperature -- 132°F (Saturated with water vapor)
Boiler efficiency -- 85.8%
System resistance -- 58.8 in wg.
(Pressurized operation)

FUEL:
Bituminous coal -- 100% through 4 mesh

TERMINALS ARE:
Inlets to: economizer, coal feeder, limestone feeder and stack
Outlets from: vacuum filter and superheater

EQUIPMENT FURNISHED:
Forced-draft fan and drive
Coal feeding system, including rate feeder, splitters and injectors
Limestone feeder
Boiler -- bottom supported (pre-assembled into two modules)
Superheater
Economizer -- extended surface
Supports for mechanical collector, economizer and precipitator
Combustion, feedwater and scrubber controls
Refractory and insulation
Mechanical collector
2-stage wet scrubber with slurry tank and pumps
Flues, ducts and dampers (not including stack)
Pneumatic transport ash-handling system
Slurry and ash dewatering system
Piping for boiler fuel system, between economizer and boiler trim piping and scrubber system connections

SERVICES FURNISHED:
Erection supervision
Erection labor

TABLE 1-34

CAPITAL COST OF COAL- AND GAS/OIL-FIRED INDUSTRIAL BOILER POLLUTION CONTROL ALTERNATIVES ($/lb/hr)

	FLUIDIZED BED COMBUSTION BOILER (packaged)	CONVENTIONAL BOILER (field erected)
COAL-FIRED(a)		
Desulfurized char	5.25	--
Low-sulfur coal	5.25	5.50
Limestone scrubbing	7.40(b)	7.60
Once-through limestone		
Limestone regeneration		
On-site facility	7.75	--
Off-site central facility	6.10	--
GAS/OIL-FIRED		
Clean fuel	4.00 - 5.00	2.44(e)
High-sulfur oil		
Once-through limestone	5.25 - 5.50(c)	3.10(f)
Wet scrubbing	7.00 - 7.50(d)	4.50

(a) Costs do not include cost of equipment before coal feeding system, e.g., railhead, bunkers. This cost is $.25 - 1.00/lb/hr.

(b) Cost is identical for two processes: calcine stone in boiler, once-through; or remove SO_2 in bed, regenerate stone, scrub effluent.

(c) Either once-through limestone or off-site regeneration.

(d) Limestone calcining - either wet scrubbing or on-site regeneration.

(e) Cost for a packaged system would be $2.00/lb/hr.

(f) The percentage sulfur removal in this case will be low.

TABLE 1-35

INDUSTRIAL BOILER COST SUMMARY

¢/10^6 Btu

	CAPITAL CHARGES[a]	FUEL COST	LABOR	SORBENT[b]	POWER	SOLIDS/SLURRY DISPOSAL[c]	TOTAL
FLUIDIZED BED COMBUSTION - COAL-FIRED							
Desulfurized char	10.95	(30)	9	---	1.1	1.90	52.96
Limestone scrubbing	15.40	30	9	2.08	1.5	13.50	71.50
Once-through dry limestone	12.70	30	9	9.69	1.1	13.50	76.00
Limestone regeneration							
On-site	16.20	30	9	1.15	1.5	5.80	63.67
Off-site	12.70	30	9	3.45	1.5	1.90	58.57
SPREADER STOKER							
Low-sulfur coal	11.50	60	9	---	1.1	1.90	83.51
High-sulfur coal - wet scrubbing	15.90	30	9	2.08	1.5	13.50	72.00
OIL/GAS-FIRED BOILER - FIELD-ERECTED, BURNING CLEAN FUEL	5.10	65	4	---	0.4	---	74.50

(a) Capital charge 16.7% per year, 70% load factor; fuel cost \$7.50/ton ≈ 30¢/10^6 BTU

(b) Raw stone \$3.00/ton; regenerated stone \$9.00/ton

(c) \$5/ton disposal cost
(cost information has varied from \$0-10/ton)

From this comparison, several projections are clearly in order:

- A fluidized bed boiler burning desulfurized char is the most economical case, if the desulfurized char is available at the cost projected, i.e., 30¢/10^6BTU.
- The limestone scrubbing scheme and once-through dry solid scheme are comparable in total cost.
- Limestone regeneration schemes have a cost advantage of approximately 15% over the other systems, if a large sulfur recovery facility is available on-site or off-site.
- A spreader stoker burning low-sulfur coal is more expensive than burning high-sulfur coal with wet scrubbing for SO_2 removal because of the high cost of low-sulfur coal.
- Although the capital cost of the field-erected oil/gas-fired boiler is substantially less than other systems, the total cost is comparable to other systems, again, because of the high fuel cost.

It is significant that the total cost of the boiler is highly sensitive to the fuel cost. Thus, the relative advantages of the alternative systems will very much depend on the future fuel situation.

Effect of Operating Variables on Cost and Design of Industrial Boiler

The cost and design of a fluid bed boiler depend mainly on operating variables such as fluidizing velocity, feed particle size, operating pressure, bed depth, bed temperature, excess air and air preheat temperature. In setting the operating variables, not only the cost, but also the efficiency of pollutants removal must be taken into consideration. An optimal boiler design will meet the requirement of air pollution abatement at a minimum cost. An optimal choice of the operating variables as such will depend on the total optimization of the operating variables involved. This is a meaningless task without accurate knowledge of the change of heat transfer coefficient in the bed, SO_2 removal efficiency, combustion efficiency, elutriation, attrition, and erosion with respect to change of different operating variables. Unfortunately, theoretical or experimental correlations which give accurate predictions are still not available. Nevertheless, the total cost of a base design may be perturbed to evaluate the effect of operating variables.

Variables capable of significantly affecting the cost and design of industrial fluid bed boilers were analyzed. The design bases of bed temperature, excess air, and air preheat temperature are usually restricted by considerations of SO_2 removal efficiency, turn-down requirement, combustion efficiency, and/or overall boiler efficiency. The scope of change of these variables is limited, as is their effect on the cost of the boiler design. The choice of bed depth dictates the fan power requirements; but once the fan size is chosen, the maximum bed depth allowable is fixed. The only two variables which affect significantly the cost and design of an industrial boiler are fluidizing velocity and operating pressure.

Fluidizing Velocity

With other design variables held constant, the bed area required for a specific fluidized bed combustor is inversely proportional to the superficial fluidizing velocity. If one coal feed point is required for a finite bed area, the total coal feed points -- and thus the complexity of the coal feed system required -- will also be inversely proportional to the superficial velocity. To minimize the elutriation of bed material and coal, larger particles have to be used at higher fluidizing velocities; this decreases the crushing cost. A higher heat transfer coefficient in the bed may also be expected, due to more agitation of solids at higher velocities. However, this will probably be partially offset by the increase of particle size, because the heat transfer coefficient in the bed is inversely proportional to the bed particle size.

Increase in fluidizing velocity tends to increase the heat-releasing rate per unit bed area of boiler. Thus, more heat transfer surface can be located in the bed, where a higher heat transfer coefficient is possible. However, there is a limit to how much heat transfer surface can be put into a unit bed volume before generating adverse effects on the coal distribution or quality of fluidization. At some point an increase in bed depth will be necessary to accommodate all the heat transfer surface in the bed; such an increase in bed depth requires high operating fan power.

In the case of the atmospheric industrial boiler, the shop-fabricated and railroad-transportable boiler is highly desirable because of its lower cost, compared to that of the field-erected boiler. In this respect, the selection of fluidizing velocity for design is very important because it dictates the size of the boiler and thus determined the possibility of shop fabrication. To evaluate the effect of fluidizing velocity on the total cost of the boiler, the change in boiler cost over the present Westinghouse - Erie City design will be estimated.

The present Westinghouse - Erie City design for a 250,000 lb/hr fluidized bed boiler consists of two shipping modules -- a primary module with the dimensions 10 ft x 14 ft x 40 ft, and a secondary module with the dimensions 8 ft x 15 ft x 40 ft, which are within the general shipping limitations of 13 ft x 16 ft x 40 ft. The designed fluidizing velocity in the primary beds is 12.5 ft/sec. Changing the design value of fluidizing velocity should not affect greatly the design of the secondary module, which houses primarily the carbon burn-up cell, superheater, and the convection zone. However, increasing the design value of fluidizing velocity will decrease the bed area of the primary beds and also require a smaller tube pitch for horizontal tubes in the bed. That means the primary module designed at 15 ft/sec should be cheaper than that designed at 12.5 ft/sec. But higher fluidizing velocity gives higher elutriation, attrition, and carbon carry-over. Increase of fluidizing velocity to more than 15 ft/sec is probably not practical at the present time.

Decreasing the design value of fluidizing velocity will necessitate an increase in bed area of the primary beds. It can be estimated that the largest shippable primary module can be designed at a fluidizing velocity of $\sim$10 ft/sec.

That means any further decrease in fluidizing velocity will require that the primary module be erected on-site. Cost investigation shows that a change of fluidizing velocity between 10 to 15 ft/sec changes the total boiler cost from the base case only <5% because only the primary module ($\sim$25% of total boiler cost) is affected. Cost analysis for a boiler designed at 8 ft/sec which requires field erection of the primary module gave a boiler capital cost of \$8.60/lb/hr, compared to \$6.10/lb/hr for the base case -- an increase of about 40%. This clearly indicates that the major saving by designing at a higher fluidizing velocity in the industrial boiler comes from the difference in cost between shop fabrication and field erection, and not from savings in bed area and the coal feed system. Thus, it is advantageous to design the industrial fluidized bed boiler at the highest fluidizing velocity feasible.

The present Westinghouse - Erie City industrial boiler design is rather conservative. Evaluation of different arrangements for heat transfer surface showed that the 250,000 lb/hr industrial boiler can actually be housed in a single module [108]. This is accomplished by installing the superheater and the convection section in the freeboard of the primary module. If the boiler can be designed at a higher fluidizing velocity ($\sim$15 ft/sec) and the heat transfer coefficient in the bed can be improved ($>$50 Btu/ft^2-hr-°F), the economizer can also be situated in the freeboard area. This will reduce the capital cost substantially. The potential of reducing the capital cost below the figures presented here is excellent; however, this requires further experimental evidence.

Pressurized Industrial Fluidized Bed Boiler

Pressurized operation requires less fluid bed area, permits operation of a deeper bed, and allows immersion of more heat transfer surface in the bed, where a higher heat transfer coefficient is possible, and thus decreases the total heat transfer surface requirement. Operation at higher pressure in deeper beds may also increase combustion efficiency and SO_2 removal efficiency because of increased residence time of coal and gas in the bed. However, gas turbines are required to recover the energy stored in high-pressure flue gas. Although this permits operation of a higher-efficiency gas-stream combined cycle, it may also require stringent particulate removal before the gas turbines and may lead to erosion, corrosion, and long-term maintenance problems of turbine blades. High-temperature and high-pressure piping, valving, ducting, particulate removal equipment, and pneumatic solid transporting lines are inherently more expensive. Hence, a potential pressurized application in utility boilers does not necessarily dictate its application in industrial boilers, because of differences in scope of design and plant economics.

For pressurized operation, a boiler enclosed in a cylindrical shell is preferred on the basis of structural strength considerations. The cylindrical vessel can be placed horizontally or vertically, and this configuration affects the design of the boiler significantly. A vertical arrangement precludes natural circulation because there is no simple way that the steam drum can be accommodated and still make the boiler railroad-shippable. A steam drum can conveniently be put inside a horizontal shell; however, the bed depth would then be limited by the diameter of the vessel, which cannot exceed 13 feet because of shipping restraints, and the advantages of the pressurized deep-bed operation are completely lost. A vertical configuration permits use of a deeper bed. For small industrial boilers, a single deep bed would be adequate; however, in this case turn-down becomes a problem. Control of bed temperature alone would provide only about 30% turn-down of total boiler capacity before SO_2 removal in the bed is greatly affected. The additional boiler turn-down would have to be provided through change of bed depth, which is complicated and expensive. Additional problems for the pressurized industrial boiler are:

- More stringent particulate removal is needed before gas turbines
- High-temperature duct from boiler to gas turbines is expensive. High-pressure coal and limestone feeding systems have to be used
- For a vertical configuration, only once-through forced circulation can be used.

All these necessities are inherently expensive compared to the atmospheric design.

A small pressurized fluid bed industrial boiler of capacity 250,000 lb/hr steam will thus not be economical compared to the present atmospheric industrial boiler. However, the economic potential of the pressurized industrial boiler increases with increases in capacity. A boiler capacity of ~50 MW will probably be economical because design and economic analysis of a four module, 300 MW, pressurized utility boiler, each module of capacity 75 MW, showed considerable savings over conventional coal-fired boilers.

Recommendations

Fluidized boilers for industrial application do not have significant overall cost and performance advantages over conventional coal-fired boilers with pollution control. On the basis of the market projections for clean fuel for industrial boilers and the technical assessment, we recommend that

- Further development work on industrial fluidized bed boilers not be carried out at present
- Fuel usage for industrial boilers be monitored to determine any change from the use of clean fuels
- Development of an industrial fluidized bed boiler be reassessed if low-sulfur char or alternate low-sulfur, low-grade fuels become available.

ELECTRIC UTILITY APPLICATION

Introduction

Concepts for utility type fluidized bed boilers evolved from the fluidized bed boiler technology which Westinghouse compiled. Foster Wheeler, using the preferred concepts, then made preliminary designs of utility boilers in the capacity range of 300 to 600 MW. Boiler subsystem costs were estimated over this size range.

Using the Ⓦ-Foster Wheeler boiler design, United Engineers and Constructors made layouts of a power plant and estimated the cost of generating energy in this plant.

Pressurized Fluid Bed Boiler System

System Concept

The system concept for the high-pressure fluidized bed boiler in a combined cycle plant is shown schematically in Figure 1-77. Combustion air is supplied to the bed by the gas turbine compressor. Coal is crushed and dried prior to being fed into the bed, the material of which is a sorbent which removes the sulfur from the products of combustion as the coal is burned. The bed is maintained at a temperature below the agglomeration point of the ash by heat transfer to the steam cycle. The spent dolomite is regenerated and recycled; and make-up dolomite, to replace that blown down, is fed in with the coal.

All of the ash from the coal is elutriated from the bed, along with a significant amount of unburned carbon and attrited sorbent. A high percentage of the particulates from the primary bed is collected in the primary separation and fed to a fluidized bed carbon burn-up cell which operates at a higher temperature and with a greater percentage of excess air than the primary bed. The effluents from the primary separator and the carbon burn-up cell are blended and passed through a secondary particulate removal device to reduce the solids content to the level required for attaining a low erosion rate in the gas turbine. The exhaust gases are cooled, after expansion through the gas turbine, by two stages of stack gas coolers. A third-stage particulate removal device may be necessary to meet stack emission standards.

The steam cycle has one stage of reheat and regenerative feedwater heating which is partially in parallel with the second-stage stack gas cooler.

The sulfur released in the sorbent regeneration is converted to elemental sulfur in a subsequent process.

Specifications

Specifications for the power plant and the fluidized bed boiler were prepared by Westinghouse for Foster Wheeler to use in making a preliminary design of the high-pressure boiler. These were as follows:

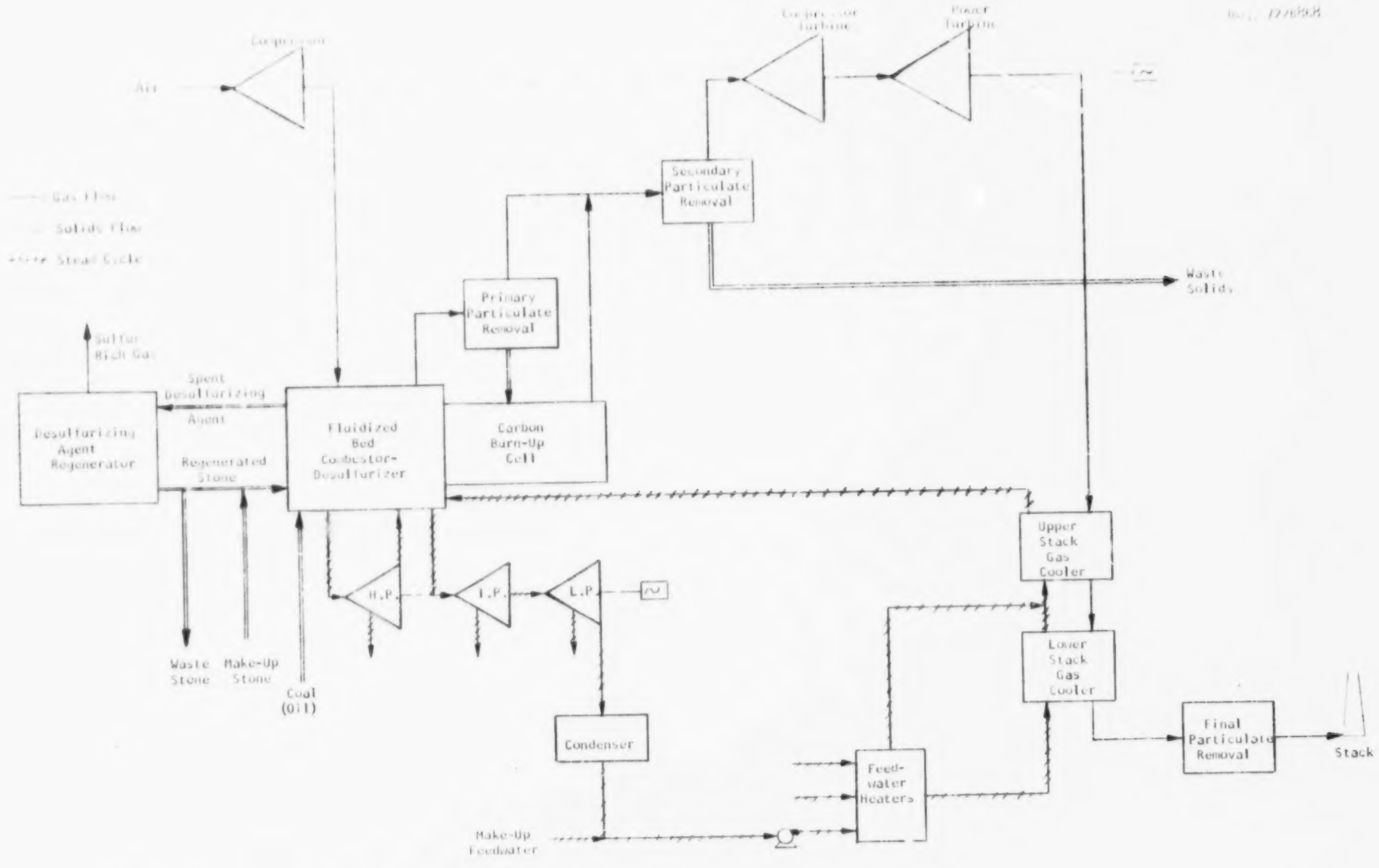

Fig. 1-77 – Pressurized fluid bed boiler power system

Plant

- Capacity - 300 MW
- Type - intermediate load
- Turn-down requirements - 4:1
- Transient requirements - 5% per min.
- Geographic location - Eastern U. S.
- Pollution control
  - Particulate emissions - 0.01 gr/SCF
  - Sulfur removal - 95%
  - Nitrogen oxide emissions - < 0.7 lb/10^6 Btu

Boiler

- Fuel - Ohio Pittsburgh No. 8 Seam Coal
  - Proximate analysis (wt %)

Volatile matter	39.5
Fixed carbon	48.7
Ash	8.5
Moisture (as mined)	3.3
	100.0

  - Sulfur content - 4.3%
  - Particle size - 1/4 in x 0
  - Complete data on coal is given on page 167.
- Sorbent - 1337 dolomite
  - Particle size - minus 1/4 in with minimum fines
  - Feed ratio - 6 x stoichiometric
  - Elutriation from beds - 5%
  - Loss during regeneration - 10%
  - Complete data on sorbent is given on page 543.

- Bed Design Parameters
  - Primary beds
    - Temperature - 1750°F design pt.
    - Size distribution of material in bed
      - Max. 5000 μ
      - Min. 500 μ
      - Mean 2500 μ
    - Superficial velocity - 6 to 9 ft/sec
    - Excess air - 10%
    - Density - 45 lb/ft^3
    - Expanded height - 10 to 15 ft
    - Elutriated material
      - Ash - 100%
      - Carbon - 6%
      - Sorbent - 5%
      - Size Distribution - see page 127
    - CO in flue gas - < 0.5% by volume
    - Gas side heat transfer coefficients
      - In bed - 50 Btu/ft^2 hr-°F
      - Above bed - 40 Btu/ft^2 hr-°F

- Carbon Burn-Up Cell
  - Temperature - 2000°F design
    - Size distribution of material in bed
      - Max. 5000 μ
      - Min. 500 μ
      - Mean 2500 μ
    - Superficial velocity - 5.7 ft/sec
    - Excess air - at least 79%
    - Combustion efficiency - 90% design pt.
    - Density - 45 lb/ft^3
    - Expanded height - 10 to 15 ft
    - Recovery of carbon elutriated from primary beds - $\geq$ 90%

Elutriated material
  Ash - 100%
  Carbon - 10%
  Sorbent - 100% of sorbent in with recovered coal plus 3% of recirculated limestone.
  Size distribution - see page 127
Gas-side heat transfer coefficients
  Same as for primary beds
Stack gas temperature - 275°F
Air preheat temperature - gas turbine compressor outlet temperature.

Power Cycle Analysis

A computer program was written to calculate plant capacity, plant heat rate, plant cost, and energy cost for combined cycle power systems using fluidized bed boilers. Figure 1-78 is the generalized block diagram of such a system. This program provides for a wide variety of power cycle configurations, including both sub- and supercritical steam cycles.

Candidate Steam Cycles. The steam conditions for the No. 4 unit at Hammond Station of Georgia Power and Light were used for the subcritical case; those at the Homer City Station for the supercritical case. These conditions are as follows:

	Subcritical (Hammond #4)	Supercritical (Homer City)
Boiler pressure-PSIA	2400	3500
Superheater temp.-°F	1000	1000
Reheater temp.-°F	1000	1000
Condenser pressure-in. Hg	3	3
Feedwater heater stages	6	7

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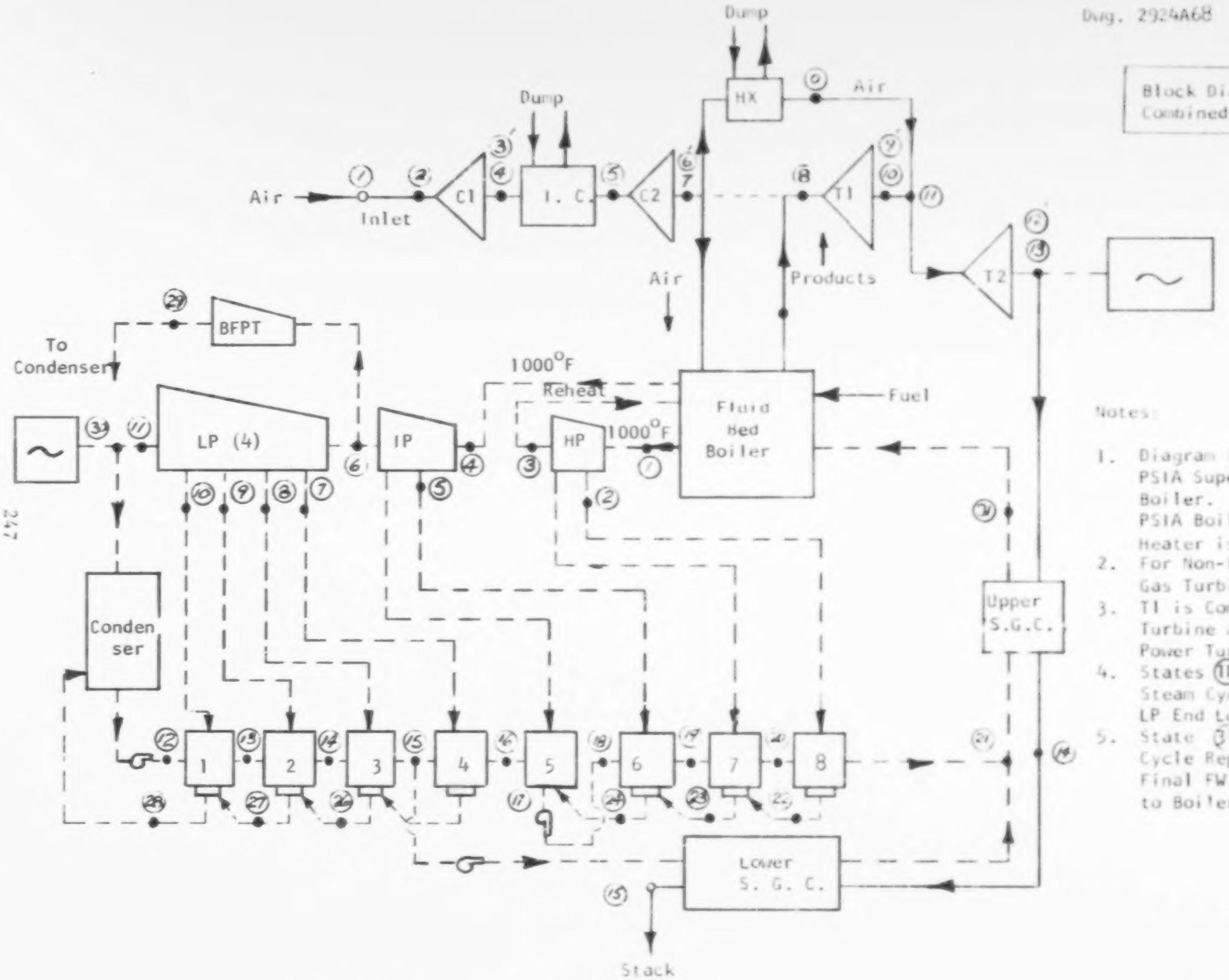
Block Diagram
Combined Cycle



Notes:

1. Diagram is for 3500 PSIA Supercritical Boiler. For 2400 PSIA Boilder, #8 Heater is Removed.
2. For Non-Intercooled Gas Turbine, ⑤=④
3. T1 is Compressor Turbine and T2 is Power Turbine
4. States ⑪ to ㉚ in Steam Cycle Represents LP End Loss.
5. State ㉛ in Steam Cycle Represents Final FW Temperature to Boiler.

Fig. 1-78—High pressure fluid bed boiler power system

In addition, the following assumptions were made for the boiler and the plant:

Bed temperature	1750°F
Turbine inlet temperature	1600°F(1)
Stack temperature	275°F
Excess air	10%
Boiler efficiency	89.6%
Auxiliary load	3%
Annual capital charge	15%
Plant capacity factor	0.8
Fuel cost	25 cents/10^6 Btu
Operating and maintenance cost (including limestone regeneration)	0.5 mills/kwh
Boiler costs: 2400 psia	$20/steam kilowatt
3500 psia	$25/steam kilowatt
Gas turbine cycle pressure drop	3%

The cost of plant components other than the boilers was obtained from a breakdown of costs for conventional plants over a capacity range of 300 to 600 MW.

Preferred Cycle Determination. First, computations were made to determine the optimum pressure ratio for this combined cycle system. Figures 1-79 through 1-82 show the results for a boiler pressure of 2400 psia and a gas turbine airflow of 650 lb/sec,(2) and Figures 1-83 through 1-86 show the results for a boiler pressure of 2400 psia and a gas turbine airflow of 1300 lb/sec. The effect of compressor intercooling was also determined for both these condition sets. These plots indicate that the best performance and the lowest cost occur with no compressor intercooling at a pressure ratio less than 10:1 for both gas turbine airflow rates. However, since the typical large

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(1) It is assumed that the internal and external heat losses from the hot gas would result in a difference of 150°F between the bed temperature and the turbine inlet temperature.

(2) This is the airflow of the Westinghouse W501 gas turbine.

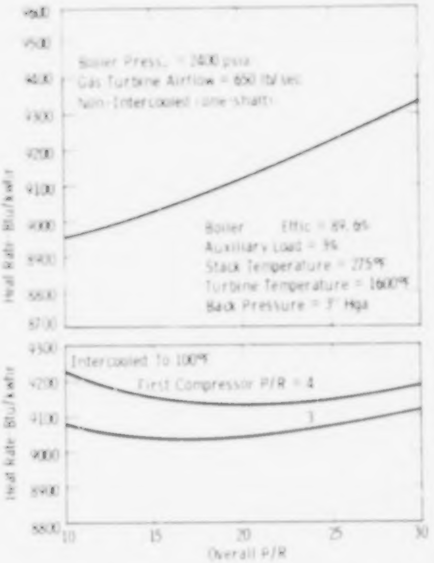


Figure 1-79–Effect of first compressor pressure ratio and overall pressure ratio on net plant heat rate

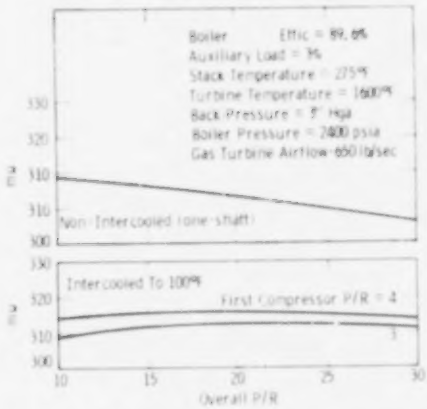


Figure 1-80–Effect of first compressor pressure ratio and overall pressure ratio on plant output

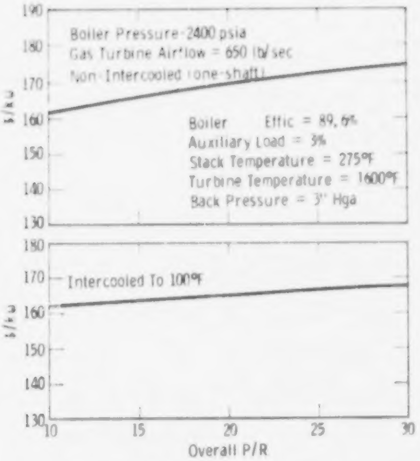


Figure 1-81–Effect of first compressor pressure ratio and overall pressure ratio on plant cost

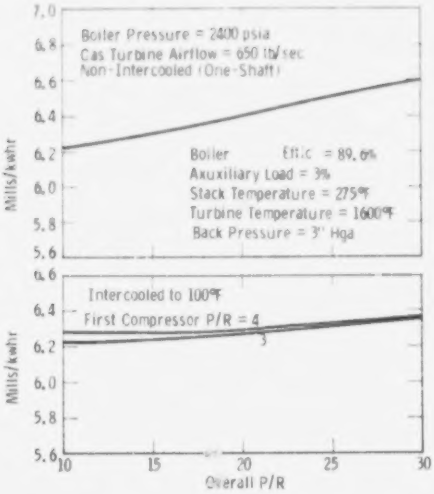


Figure 1-82 –Effect of first compressor pressure ratio and overall pressure ratio on energy cost

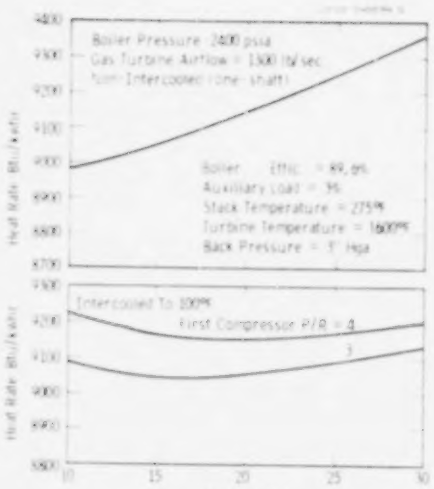


Figure 1-83—Effect of first compressor pressure ratio and overall pressure ratio on net plant heat rate

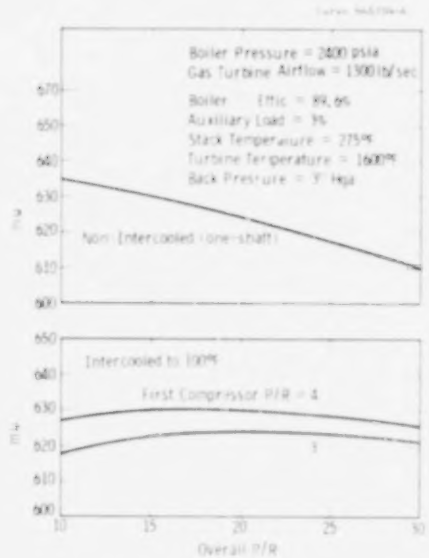


Figure 1-84—Effect of first compressor pressure ratio and overall pressure ratio on plant output

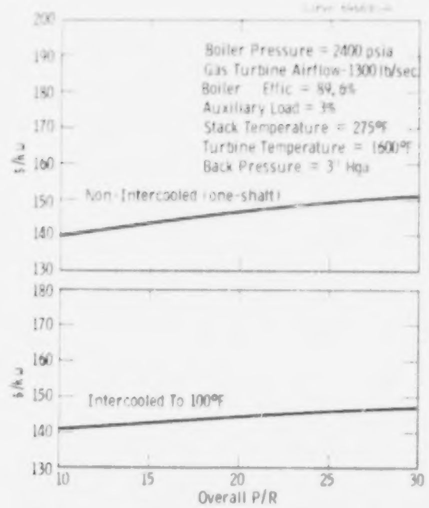


Figure 1-85—Effect of first compressor pressure ratio and overall pressure ratio on plant cost

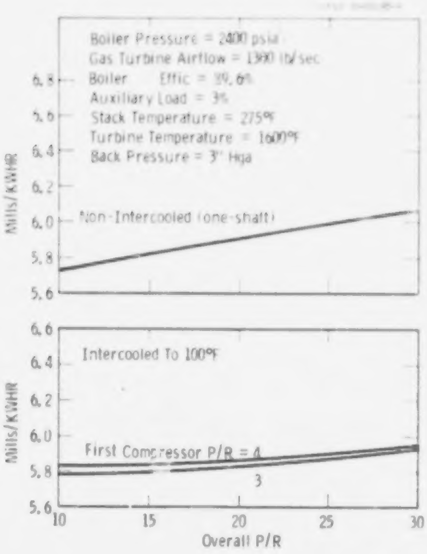


Figure 1-86—Effect of first compressor pressure ratio and overall pressure ratio on energy cost

utility type gas turbine has a pressure ratio of about 10:1, this is taken to be the minimum practical value.

Figures 1-87 through 1-94 show the results of the supercritical steam conditions and gas turbine airflows of 650 and 1300 lb/sec. Here, again, the performance and costs are best at an overall pressure ratio of 10:1 with no compressor intercooling. Comparing the results for 3500 psia with those for 2400 psia shows that the heat rate is lower and the power higher for supercritical steam conditions than for subcritical conditions. However, the specific plant cost and the energy cost are higher for supercritical than for subcritical conditions.

In view of the above, it was decided that the optimum conditions for use in making the preliminary design of the combined cycle plant with a fluidized bed boiler are:

- Simple cycle gas turbine with a pressure ratio of 10:1 and without compressor intercooling
- Subcritical (2400 psia) steam conditions for both 318 and 635 MW level.

Design Point Conditions. The design point conditions for the combined cycle plant with the fluidized bed boiler are as follows:

	CAPACITY - MW	
	318	635
Boiler pressure-psia	2400	2400
Initial steam flow-lb/hr	1,727,020	3,454,040
Steam temperature-°F		
Initial	1000	1000
Reheat	1000	1000
Condenser pressure-in.Hg	3	3
Final FW temperature-°F	578	578
Compressor airflow-lb/sec	650	1300
Gas turbine temperature-°F	1600	1600

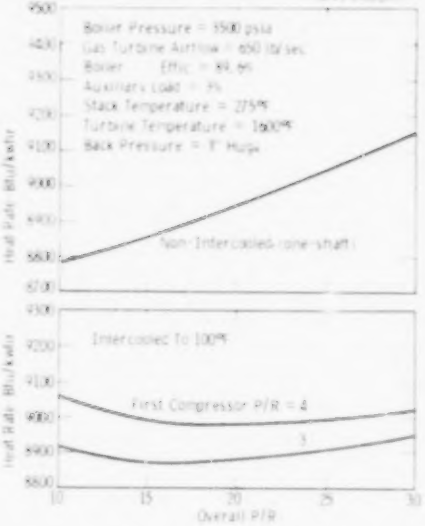


Figure 1-87—Effect of first compressor pressure ratio and overall pressure ratio on net plant heat rate

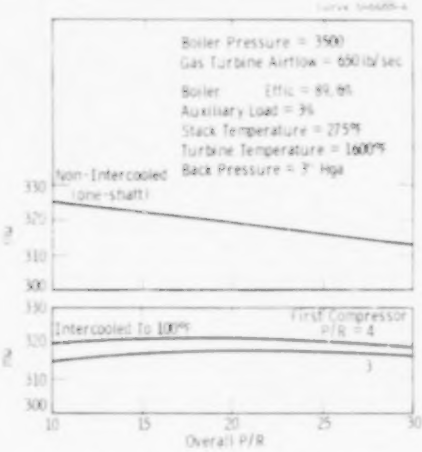


Figure 1-88—Effect of first compressor pressure ratio and overall pressure ratio on plant output

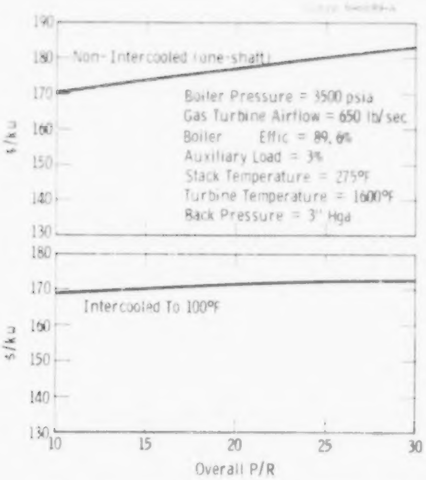


Figure 1-89—Effect of first compressor pressure ratio and overall pressure ratio on plant cost

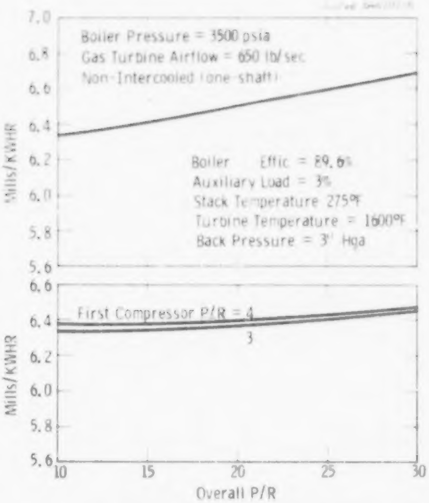


Figure 1-90—Effect of first compressor pressure ratio and overall pressure ratio on energy cost

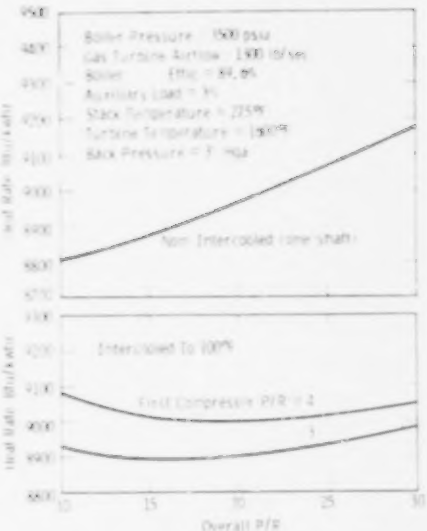


Figure 1-91—Effect of first compressor pressure ratio and overall pressure ratio on net plant heat rate

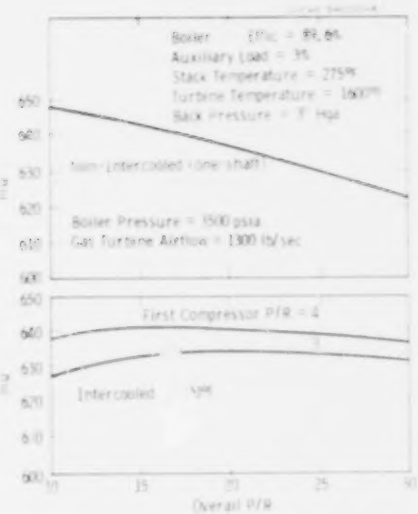


Figure 1-92—Effect of first compressor pressure ratio and overall pressure ratio on plant output

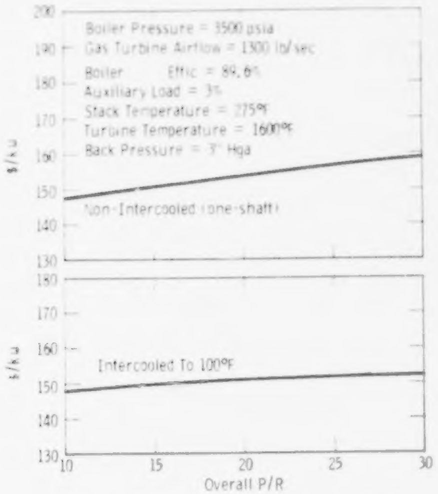


Figure 1-93—Effect of first compressor pressure ratio and overall pressure ratio on plant cost

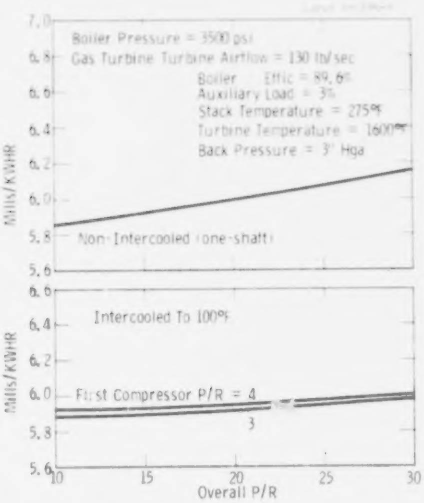


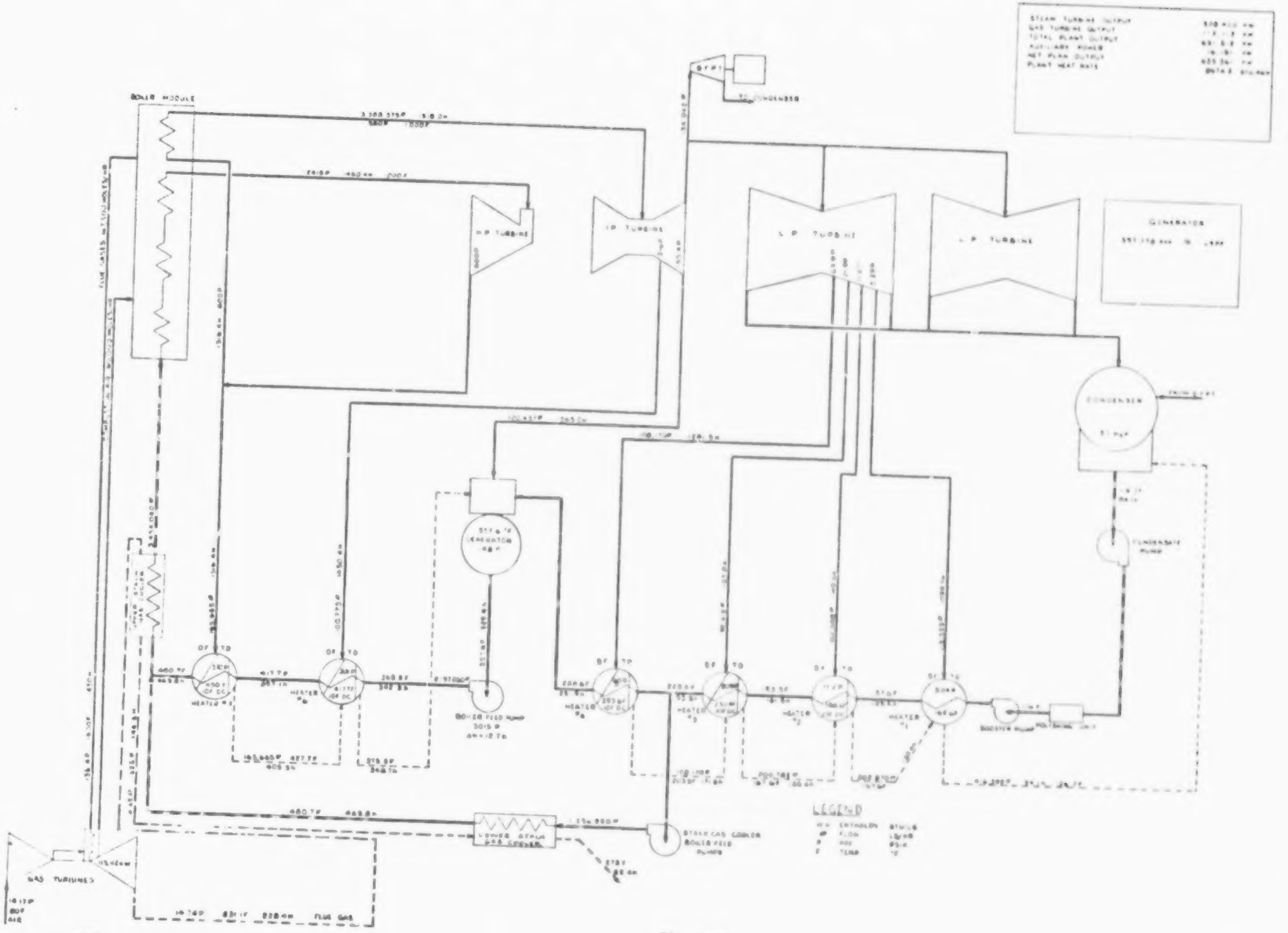
Figure 1-94—Effect of first compressor pressure ratio and overall pressure ratio on energy cost

	CAPACITY - MW	
	318	635
Gas turbine pressure ratio	10:1	10:1
Turbine cooling air	5%	5%
Gas turbine pressure drop	3%	3%
Excess air	10%	10%
Stack temperature-°F	275	275
Gas turbine PWR-KW	56,557	113,113
Steam turbine PWR-KW	270,029	548,400
Net plant power-KW	318,485	635,362
Plant heat rate-Btu/kwh	8953	8974

Appendix J , prepared by United Engineers and Constructors, contains a flow diagram of the combined cycle plant. It is reproduced in Figure 1-95. The design point thermodynamic data are given at each station on this diagram.

Gas Turbine Pressure Loss Effect. In calculating the plant performance for the above design points, the pressure loss between the gas turbine compressor outlet and turbine inlet was assumed to be 3%. Since many system components contribute to this pressure loss, and in most cases there are trade-offs between component cost and pressure drop, a sensitivity analysis was made of the effect of the gas turbine pressure losses on the plant capacity and heat rate. The results of this analysis are shown in Table 1-36 over a range of pressure losses from 3% to 10%. This shows that a change of one percentage point in pressure loss between the gas turbine compressor exit and turbine inlet causes a 0.1% decrease in plant power and a 0.1% increase in plant heat rate. The reasons for the relatively small effect of gas turbine pressure loss on plant capacity and heat rate are:

- Only the gas turbine is directly affected, and it contributes a minor fraction of the plant power
- Decrease in gas turbine power results in higher gas turbine exit temperature and increased steam power.



485 MW FLUIDIZED BED
BOILER COMBINED CYCLE PLANT
PERFORMANCE DIAGRAM

2551

Fig. 1-95
2.5.5.2

255.3

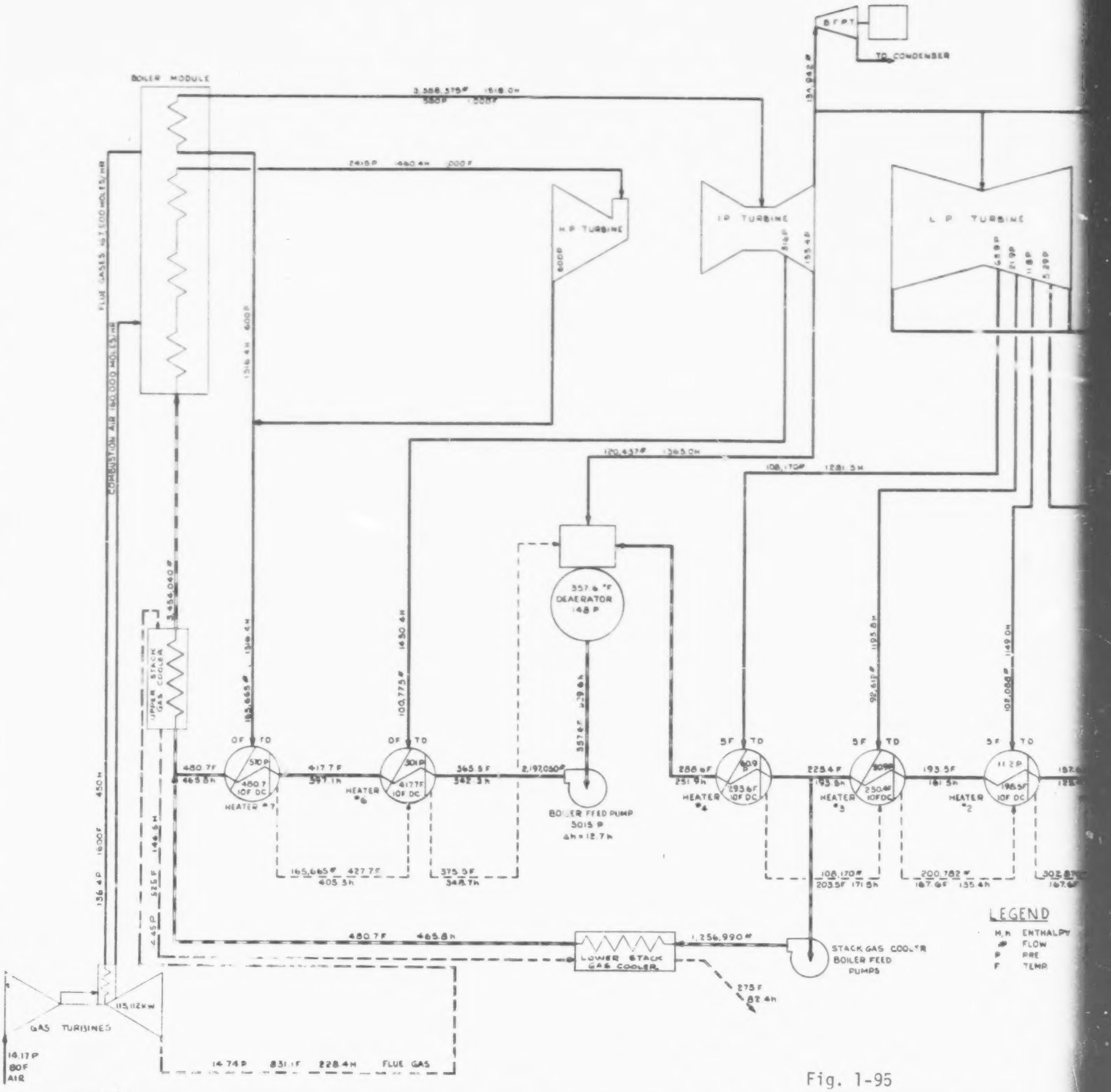


Fig. 1-95

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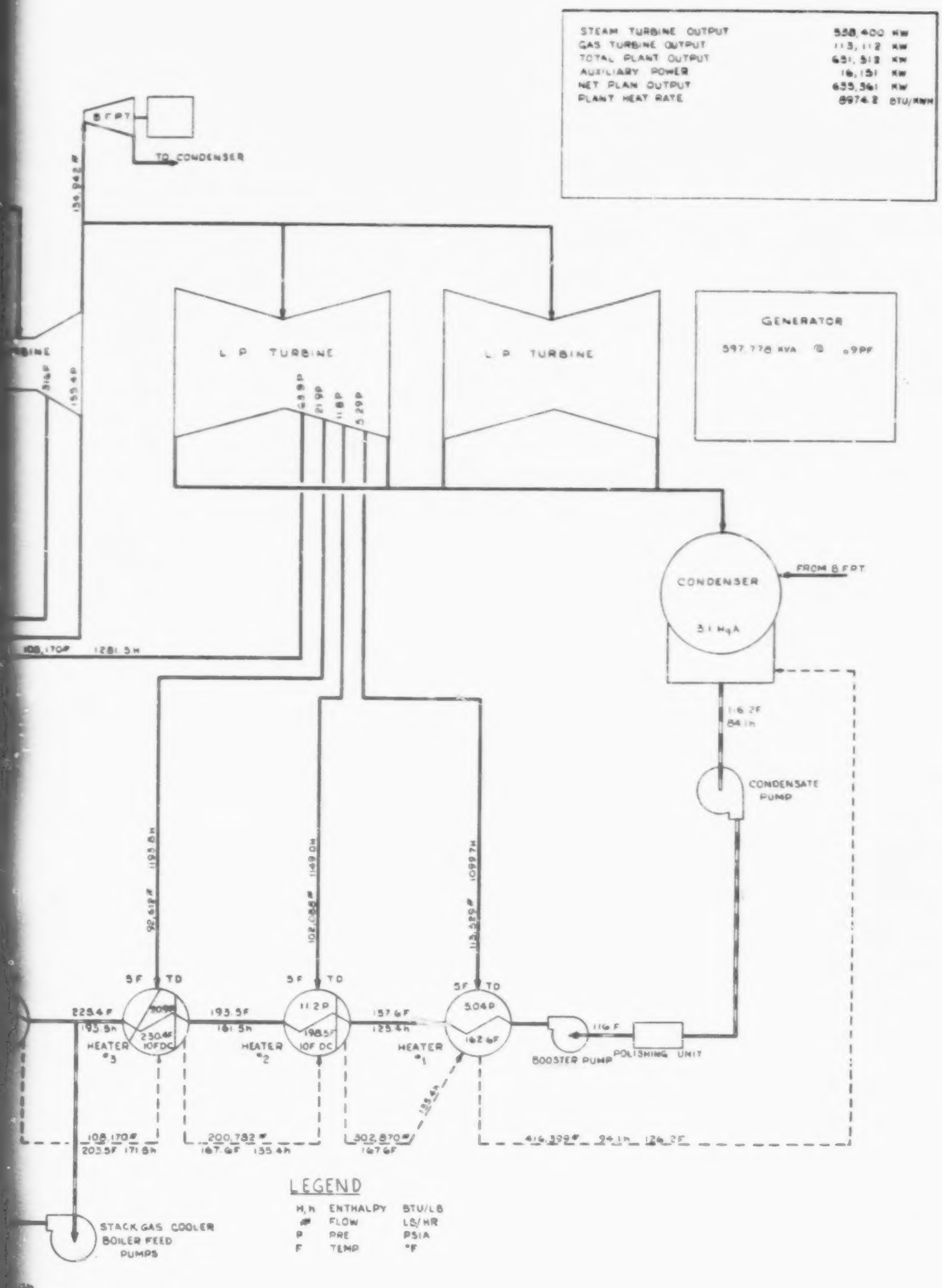
255.2



635 MW FLUIDIZED BED
BOILER COMBINED CYCLE PLANT
PERFORMANCE DIAGRAM
united engineers

Fig. 1-95

255.2

Table 1-36

EFFECT OF GAS TURBINE PRESSURE DROP ON DESIGN POINT PLANT PERFORMANCE

PRESSURE DROP[(a)]	GAS TURBINE POWER -- MW	STEAM TURBINE POWER -- MW	NET PLANT POWER -- MW	PLANT HEAT RATE -- BTU/KW HR
0.03	113.1	538.4	635.4	8975
0.04	112.1	538.9	634.8	8983
0.05	111.0	539.4	634.2	8992
0.06	109.9	539.8	633.6	9000
0.07	108.8	540.3	633.0	9009
0.08	107.7	540.3	632.3	9018
0.09	106.6	541.3	631.7	9028
0.10	105.5	541.8	631.0	9037

(a) Pressure loss between gas turbine compressor outlet and turbine inlet divided by compressor outlet pressure.

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Condenser Pressure Effect. A steam condenser pressure of 3 in. Hg was assumed for the design point calculations. This is typical for a power plant with a cooling tower; a pressure of 1-1/2 in. Hg is typical for a plant with heat rejection to a body of water. Since cooling water temperature is a major variable among plant sites, design point calculations were made for the anticipated range of condenser pressures. These are shown in Table 1-37.

Part Load Performance. The turn-down requirements for fluid bed combined cycle power plant have been specified as 75% of design load. As in any power system, the output of the combined cycle plant with fluidized bed boilers is roughly proportionate to the fuel input. In many combustion systems efficient burning requires a nearly constant fuel-air ratio, so the air rate must be maintained approximately proportionate to fuel rate. In fluidized bed combustion, however, good burning can be obtained over a wide range of fuel-air ratios.

Another possible constraint on air rate is that imposed by the hydrodynamic requirements of the bed. Figure 1-51 shows that the ratio of maximum to minimum superficial velocities for a given bed material is of the order of 10:1. Since there are no severe constraints on either fuel-air ratio or airflow rate in fluidized bed combustion, these two variables are effective parameters for controlling energy input to the fluidized bed boiler.

The steam cycle power output is primarily a function of the heat transferred to the working fluid in the fluidized bed boiler, i.e.,

$$q = \overline{U}\ S\ \overline{\Delta T}_{LM} \qquad (1\text{-}32)$$

where q = total heat transferred to the working fluid-Btu/hr

$\overline{U}$ = area weighted average overall heat transfer coefficient-Btu/(hr)(ft^2)(°F)

S = total heat transfer surface in the fluidized bed boiler both in and out of the bed-ft^2

$\overline{\Delta T}_{LM}$ = area weighted average log mean temperature difference between the bed or hot gas and the working fluid-°F.

TABLE 1-37

EFFECT OF CONDENSER PRESSURE ON DESIGN POINT PLANT PERFORMANCE

CONDENSER PRESSURE - in Hg	1	2	3
Boiler pressure--price	2,400	2,400	2,400
Steam temperature--°F			
Initial	1,000	1,000	1,000
Reheat	1,000	1,000	1,000
Compressor airflow--lb/sec	1,300	1,300	1,300
Gas turbine temperature--°F	1,600	1,600	1,600
Gas turbine pressure ratio	10:1	10:1	10:1
Gas turbine pressure loss--%	3	3	3
Gas turbine cooling air--%	5	5	5
Excess air--%	10	10	10
Initial steam flow--lb/hr	3,454,040	3,454,040	3,454,040
Final feedwater temp.--°F	578	578	578
Gas turbine power--kw	113,114	113,114	113,113
Steam turbine power--kw	575,217	553,350	538,400
Net plant power--kw	671,075	649,864	635,362
Plant heat rate--Btu/kw hr	8,497	8,775	8,974

The overall heat transfer coefficient decreases slightly with decreasing steam flow because of Reynolds number effects. The temperature of the working fluid in the Rankine cycle is fixed with respect to plant load for both constant pressure and variable pressure systems. Therefore, for a fixed heat transfer surface area, the only independent parameter for controlling steam cycle power is bed temperature.

There are definite limitations on the operating temperature of a fluidized bed when burning coal. The maximum value of ~1750°F is set by the ash softening characteristics, and the minimum temperature, about 1600°F for limestone and 1400°F for dolomite, is imposed by the reaction kinetics of the sulfur sorbent.

When the range of variability of the bed temperature is limited, the capacity for turn-down of steam power in a fluidized bed boiler can be extended by varying the amount of heat transfer surface immersed in the bed. This can best be accomplished by reducing the bed depth and uncovering a part of the heat transfer surface, exposing it to the hot gas effluent from the bed, from which it absorbs heat. However, the heat transfer coefficient is significantly lower for surface in the overhead than for surface in the bed, so there is a change in the area weighted average heat transfer coefficient.

The power output of the gas turbine is a function of the turbine inlet temperature, the compressor pressure ratio, and the compressor airflow for fixed component efficiencies, i.e.,

$$P_{GT} = \phi(Tg, PRc, Wa). \tag{1-33}$$

For a given turbine configuration, the turbine flow number $(\frac{Wg\sqrt{Tg}}{P})$ is constant and PRc = ϕ (Wa, Tg). Therefore, for a fixed gas turbine configuration

$$P_{GT} = \phi(Tg, Wa). \tag{1-34}$$

The gas temperature out of the fluid bed boiler is a function of the bed temperature, which makes the gas turbine power output a function of the bed temperature and the compressor airflow, i.e.,

$$P_{GT} = \phi(T_b, Wa). \qquad (1\text{-}35)$$

The axial flow compressor used in gas turbines has the pressure ratio versus flow rate characteristics shown in Figure 1-96. For a fixed speed, the flow rate is essentially constant over a wide range of pressure ratios. Airflow, however, can be reduced by approximately 1/3 by the use of variable inlet guide vanes.

It has been shown that the energy input to the boiler, the steam cycle power, and the gas turbine power in the combined cycle fluidized bed boiler are a function of the following four parameters:

- Airflow rate
- Fuel-air ratio
- Bed temperature
- Bed depth.

There are three possible modes of load control for the combined cycle fluidized bed boiler using different combinations of these parameters:

Mode 1

- Constant fuel-air ratio and bed temperature
- Variable airflow and bed depth
- Fuel rate decreases to maintain constant fuel-air ratio
- Bed depth is reduced simultaneously with airflow to maintain constant bed temperature; turbine inlet pressure decreases to satisfy constant turbine flow number
- Steam power decreases due to reduced U; gas turbine power decreases due to lower Wa.

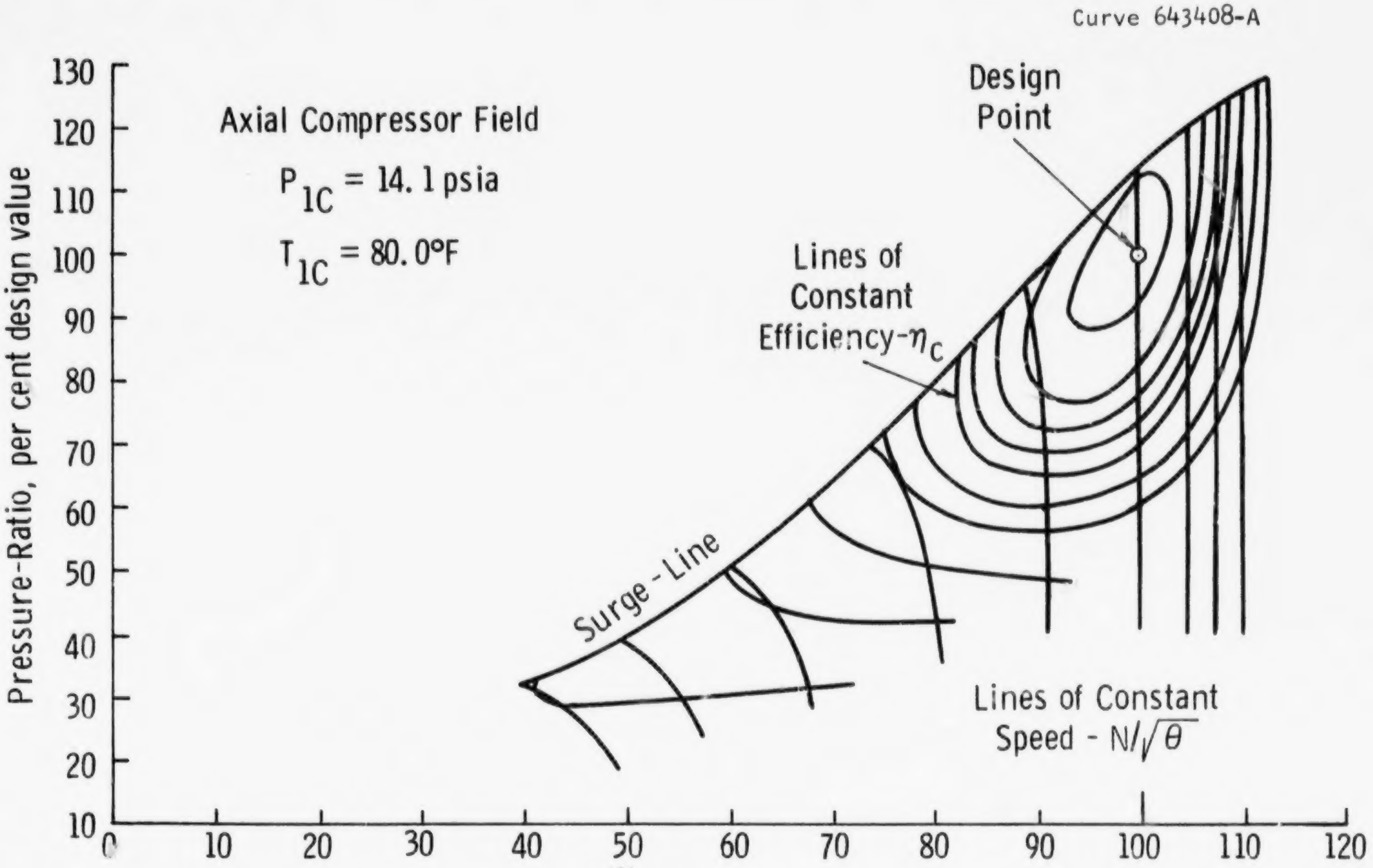


Fig. 1-96-- Discharge-Flow $\frac{W_a\sqrt{\theta}}{\delta}$ per cent design value

Mode 2

- Constant Wa and bed temperature
- Variable bed depth and fuel-air ratio
- Bed depth is reduced simultaneously with fuel-air ratio to maintain constant bed temperature
- Turbine inlet pressure decreases slightly to satisfy constant turbine flow number with reduced fuel flow
- Steam power decreases due to reduced U; gas turbine power would be nearly constant.

Mode 3

- Constant bed depth and airflow
- Variable fuel-air ratio and bed temperature
- Gas turbine inlet pressure reduces to satisfy constant turbine flow number
- Steam power decreases due to lower bed temperature
- Gas turbine power decreases due to lower gas temperature.

The Westinghouse W501 gas turbine has a design airflow of 650 lb/sec. By using variable inlet guide vanes, the airflow can be reduced by 1/3 to 433.5 lb/sec. Computations have shown that when load control mode 1 is used, the system output from this combined-cycle plant is nearly proportional to airflow. Using mode 2 sequentially with mode 1 can reduce the system output to as low as 45% of the design value. This indicates that the use of mode 2 alone can effect a power reduction of as much as 34%. The performance data for part-load operation using modes 1 and 2 in sequence are given in Table 1-38. The part-load heat rate for the fluid bed combined cycle system with operating modes 1 and 2 in sequence is plotted in Figure 1-97, along with that for a conventional coal-fired plant. This shows that there is a minimum heat rate at about 75% load which is over 1% lower than the design point heat rate.

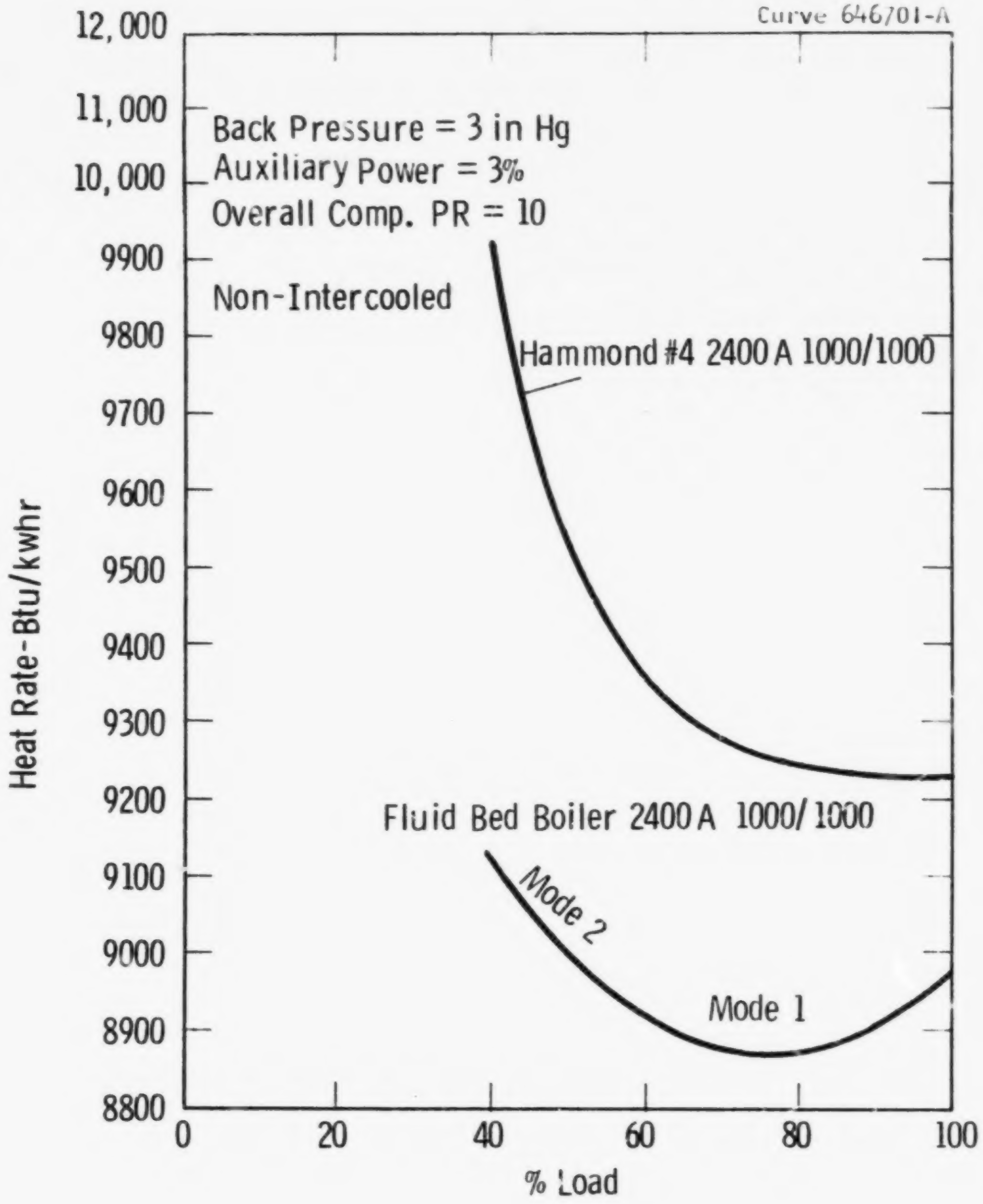


Figure 1-97—Part Load characteristics fluid bed combined cycle with operating modes 1 & 2 in sequence

TABLE 1-38

PART LOAD PERFORMANCE UNDER OPERATION MODES 1 AND 2

FRACTION STEAM POWER	STEAM FLOW LB/HR	AIRFLOW LB/SEC	FUEL AIR RATIO	EXCESS AIR %	HEAT RATE BTU/KWH	PLANT OUTPUT KILOWATTS
Max. Guarantee	3,454,040	1300	.091937	10	8974	635344
100%	3,386,314	1278.9	.091937	10	8968	625551
80%	2,709,051	1049.8	.091937	10	8873	519033
60%	2,031,788	867	.086593	16.8	8899	402533
40%	1,354,526	867	.062535	61.7	9057	285630

Two modes of load control have been considered for the high-pressure fluidized bed boiler power system wherein the bed temperature is held constant over the full range of load. One mode varies airflow keeping the fuel-air ratio constant; the other varies the fuel-air ratio, keeping airflow constant. Any mode wherein bed temperature is kept constant requires that the bed depth be varied to satisfy the heat transfer relationship.

The combined cycle performance calculation program was used to generate a matrix of performance data for a series of turbine inlet temperatures and fuel-air ratio value sets (see Table 1-39). A procedure for determining a system operating line for variable bed temperature from this matrix was developed. By definition, the system operating line is a locus of condition sets wherein the heat transferred from the bed to the steam cycle matches the power derived from the steam cycle.

For the high-pressure system, all of the heat transfer surface for the steam cycle--except for feedwater heating--is located in the fluidized bed or in the transition zone above the bed, where the heat

TABLE 1-39

PERFORMANCE DATA FOR VARIABLE BED TEMPERATURES

Steam Cycle Load Fraction (ref.)	1.02	1.00	0.8	0.6	0.4
Turbine inlet temp. - °F	1600				
Plant power - mw	635.4	627.4	543.3	446.2	325.6
Steam power - mw	538.4	530.5	446.9	350.1	229.3
Plant heat rate - Btu/kw hr	8974	8967	8869	8846	8952
Fuel-air ratio	0.0919	0.0907	0.0777	0.0636	0.0470
Turbine inlet temp. - °F	1500				
Plant power - mw	619.5	611.6	527.4	430.5	311.5
Steam power - mw	537.5	529.5	445.7	348.8	229.3
Plant heat rate - Btu/kw hr	9091	9080	8999	8998	9165
Fuel-air ratio	0.0908	0.0895	0.0765	0.0625	0.0460
Turbine inlet temp. - °F	1400				
Plant power - mw	603.5	595.6	511.5	414.9	297.3
Steam power - mw	536.1	528.2	444.1	347.2	229.3
Plant heat rate - Btu/kw hr	9211	9202	9134	9161	9392
Fuel-air ratio	0.0896	0.0884	0.0753	0.0613	0.0451

transfer coefficient is nearly as high as it is in the bed. The heat transfer equation $q_i = U_i A_i \Delta T_{LM_i}$ has to be satisfied for each bed. The overall heat transfer coefficient varies measurably among the evaporator, the superheater, and the reheater because of differences in steam-side coefficients; it also varies with load because of Reynolds number effects. Therefore:

$$q_i = U_i A_i \Delta T_{LM_i} \tag{1-36}$$

and

$$q = \Sigma_i^n q_i = \Sigma_i^n U_i A_i \Delta T_{LM_i} . \tag{1-37}$$

Also:

$$q_i = W_i (h_{i_{out}} - h_{i_{in}}) . \tag{1-38}$$

Combining 1-37 and 1-38 gives

$$A_i = \frac{W_i (h_{i_{out}} - h_{i_{in}})}{U_i \ \Delta T_{LM_i}} .$$

For constant bed depth

$$A = \Sigma_i^n A_i = C .$$

For a constant pressure steam cycle, the q_i values for a given fraction of reference steam power are constant with variable bed temperature in all beds except the one containing the pre-evaporator. Therefore, the distribution of heat absorbed by the various beds shifts with bed temperature, and for this reason the temperatures of all the beds cannot be the same at reduced load. It was assumed for this analysis that the functions of the pre-evaporator bed and the two superheater beds would be lumped and that the temperatures of these beds would be maintained the same at all loads. The temperature of the reheat bed would therefore be different from that of the other beds.

The conditions and assumptions used in this analysis are as follows:

- Temperature and pressure profile at design point (see Figure 1-98)

  t_1 = 578.3°F p_1 = 2632 psig
  t_2 = 675°F
  t_3 = 671°F
  t_4 = 1000°F p_4 = 2486 psig
  t_5 = 650°F p_5 = 601 psig
  t_6 = 1000°F p_6 = 581 psig
  t_B = 1750°F

- It has been assumed that the bed temperature is 1750°F and the gas turbine inlet temperature is 1600°F. This difference would decrease with decreasing bed temperature. It was assumed that the difference between these two temperatures would vary linearly from 150°F at a bed temperature of 1750°F to zero at a bed temperature of 550°F.

- The carbon burn-up cell performs a portion of the pre-evaporation function as shown in Figure 1-99. It was assumed that the heat absorbed by the steam cycle from the carbon burn-up cell would be a constant 5% of the total heat absorbed over the load range. If a constant temperature must be maintained in the carbon burn-up cell, the bed depth will have to be varied accordingly.

Dwg. 6169A24

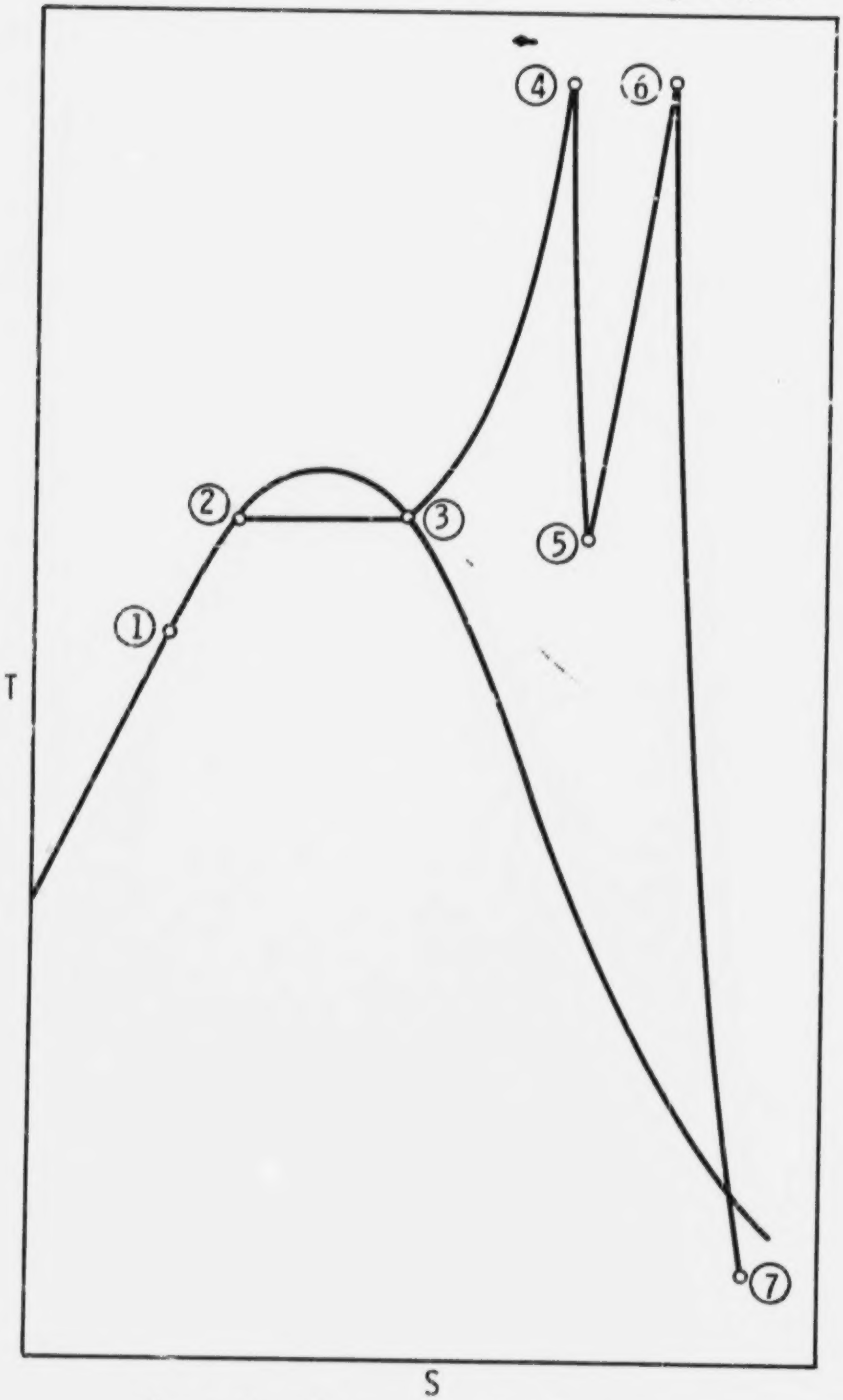


Fig. 1-98 — Steam cycle T—S diagram

Curve 645594-A

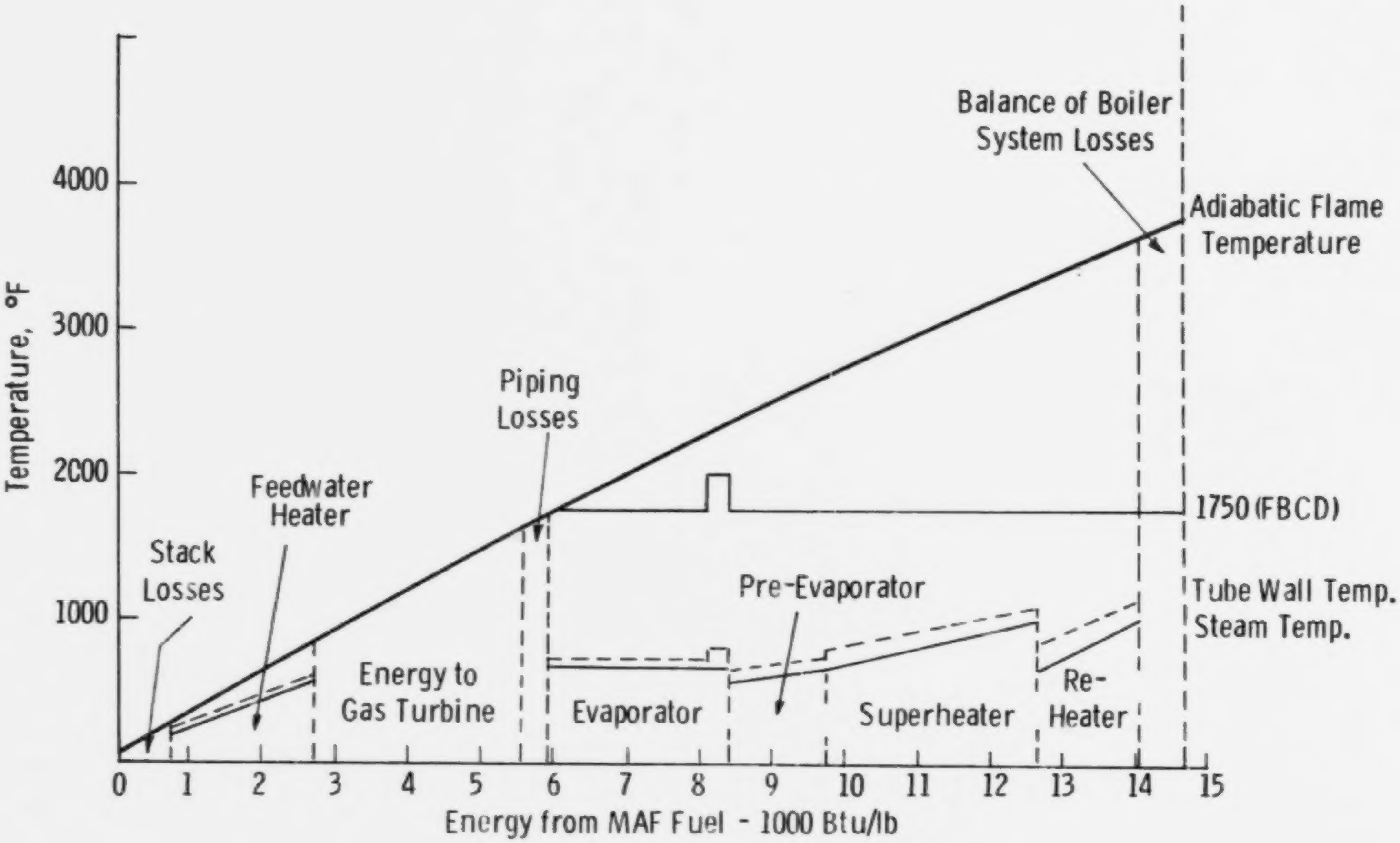


Fig. 1-99–Temperature-energy diagram for pressurized utility boiler

- It was assumed that the enclosure wall surface where evaporation takes place is exposed to the same temperature as the lumped beds.
- The feedwater temperature as a function of the fraction of steam power and the gas turbine inlet temperature is given in Figure 1-100.
- The temperature of the steam returning to the reheater is given as a function of the fraction of steam power in Figure 1-101.
- Overall Heat Transfer Coefficient

	FRACTION STEAM POWER (REF)		
	1.02(1)	0.8(2)	0.6(2)
Pre-evaporator	47	46	45
Evaporator	47	46	45
Superheater	43	42	41
Reheater	39	38	37

---

(1) Calculated by Foster Wheeler

(2) Assumed

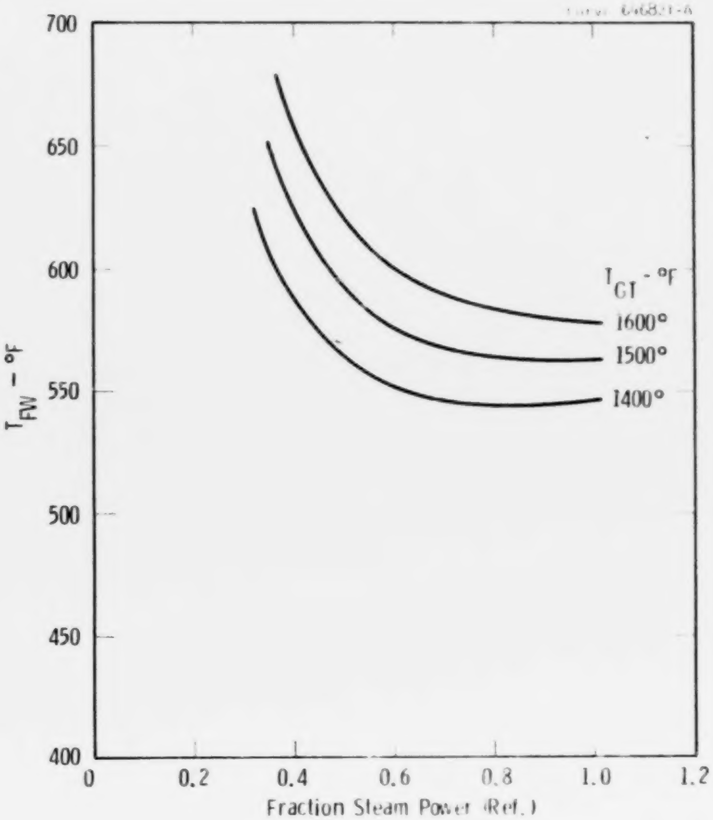


Fig. 1-100 – Feedwater temperature in high-pressure fluidized bed boiler

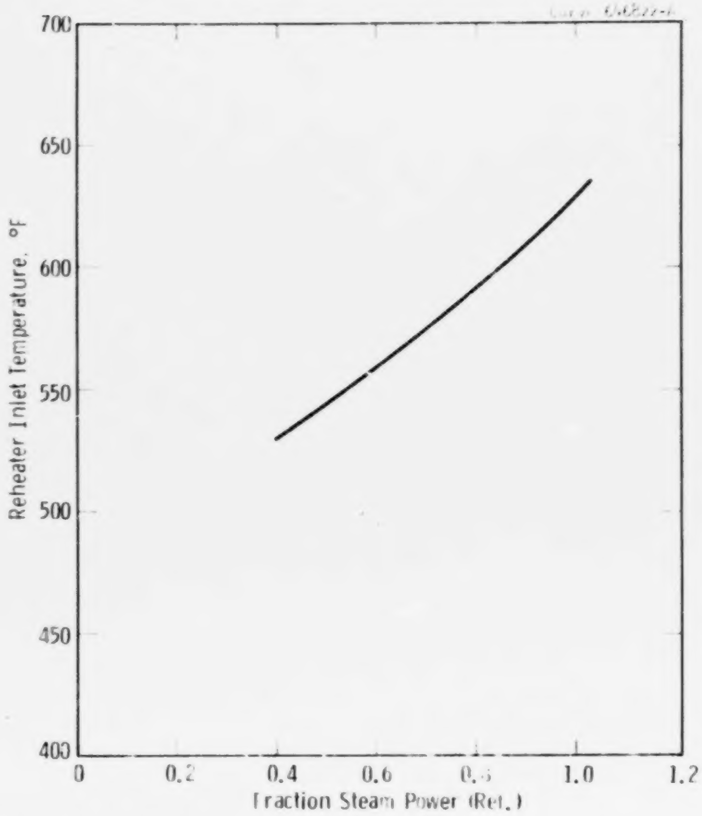


Fig. 1-101 – Temperature of steam to reheater in high-pressure fluidized bed boiler

The operating lines for the lumped beds and the reheater were determined for the above conditions and assumptions. These are shown plotted on the performance matrix in Figure 1-102. This shows that the reheater bed will operate at a higher temperature than will the other beds. The operating line for the reheat bed is shown for reference only and is valid only for bed temperature and fuel-air ratio, since it obviously can have no significance for heat rate. No correction was made for the effect of the higher temperature of the reheat bed or for the gas turbine inlet temperature.

The minimum operating bed temperature is determined by the requirements for absorption of the sulfur released from the coal by the bed material. For limestone the projected minimum bed temperature is >1600°F; for dolomite it is 1300°F to 1400°F. Figure 1-102 shows that the power for a minimum bed temperature of 1600°F is 528 MW, or 83% of design power, whereas for bed temperatures of 1300°F and 1400°F the powers are 325 MW (51%) and 395 MW (62%) respectively.

Reducing the compressor airflow by 1/3, with a constant bed height and a bed temperature of 1300°F, would not affect the steam power but would reduce the gas turbine power by about 15 MW to about 48% plant load. The part load performance line for the fluid bed combined cycle plant with operating mode 3 is also superior to that of a conventional coal-fired plant as shown in Figure 1-97, but it is not as good as for operating modes 1 and 2 in sequence.

Since varying the bed depth requires expensive materials handling equipment to avoid a large response time constant, an operating mode which uses constant bed temperature is not considered to be the optimum turn-down method. In addition, the turn-down capability with constant bed height is nearly as good as that with variable bed height. Therefore, the preferred method for controlling load in the high-pressure fluidized bed boiler is to vary the bed temperature, possibly in combination with variation in airflow.

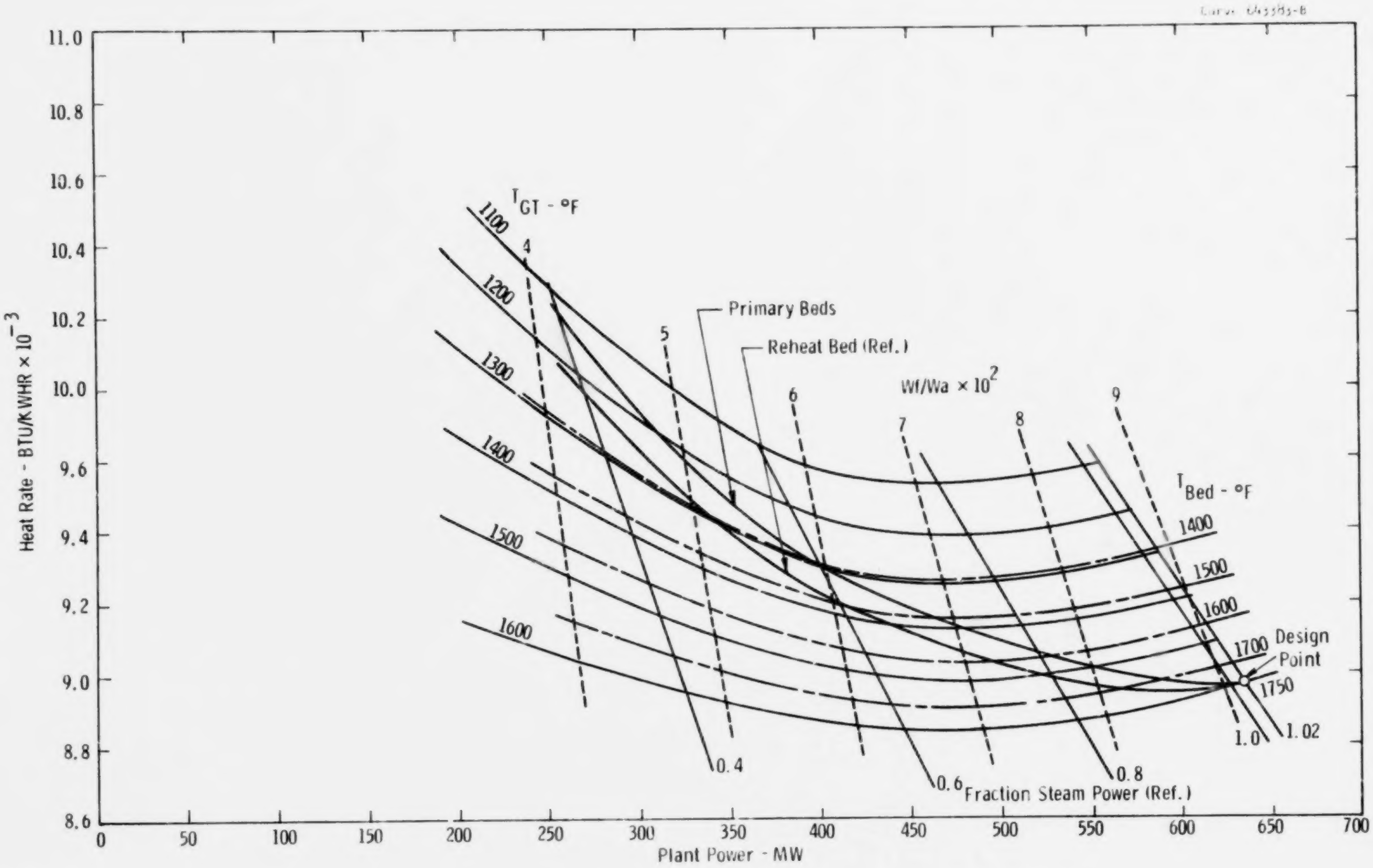


Fig. 1-102—Operating lines for high pressure fluidized bed system with variable bed temperature load control

The turn-down capability of a single fluidized bed boiler coupled with a single gas turbine is only about 50%, whereas the specified turn-down for this power plant is 4:1. Therefore, a modular plant configuration will be necessary to meet the turn-down requirements. The 318 MW plant gas turbine requirements could be satisfied by either one W501 gas turbine or two W251 gas turbines. In view of the turn-down requirements, two W251 gas turbines were selected for this size plant. Two W501 gas turbines would be required for a 635 MW plant and one or more boiler modules would be required for each gas turbine module.

The relationship between plant and module turn-down capabilities is as follows:

$$X_{P_1} = \frac{1}{n} X_m \qquad (1\text{-}39)$$

where

X_{P_1} = minimum load fraction of plant with one module operating,

n = number of modules in plant, and

X_m = minimum load fraction of module.

The values of X_m required to give a plant turn-down ratio of 4:1 are

n	X_m
1	0.25
2	0.50
3	0.75
4	1.00

In addition to being able to meet the minimum load, it is advantageous to have continuity in plant load between the minimum and full load values. Continuity is satisfied when

$$X_{Pi} = X'_{P(i-1)}$$

$$X_{Pi} = i \cdot \frac{1}{n} \cdot X_m$$

$$X'_{P(i-1)} = \frac{i-1}{n}$$

$$X_m = \frac{i-1}{i} \quad \text{for i values greater than 1}$$

$$X_m \neq \emptyset\,(n)$$

where

X_{Pi} = minimum load fraction with i modules operating and

$X'_{P(i-1)}$ = maximum load fraction with i-1 modules operating.

This shows that a module turn-down capability of 50% is required to satisfy continuity, regardless of the number of modules in the plant.

For the plant with two gas turbine modules, a module turn-down capability of 50% will satisfy the requirements for both minimum load and load continuity. The calculations indicate that this can just be met with a 1300°F temperature in the primary beds of the two boiler modules operating in unison.

If the minimum bed operating temperature for satisfactory SO_2 removal is 1400°F, the gas turbine module turn-down capability with two boiler modules operating in unison is about 60%. This does not meet either the minimum load or the continuity requirement. Operation of the gas turbine module with one of the boiler modules turned off must be considered. The idle temperature for the W251 and W501 gas turbines is about 970°F. When the primary beds in both boiler modules are at 1400°F, the gas turbine inlet temperature is about 1300°F; if one bed is turned off, the turbine inlet temperature is about 1000°F. Therefore, operating a gas turbine module with one bed turned off and the other at a primary bed temperature of 1400°F is feasible.

The power ranges attainable with all possible combinations of gas turbine and boiler modules operating between primary bed temperatures of 1300°F and 1750°F are plotted in Figure 1-103. This shows that a discontinuity exists between 27% and 31% of design power when the minimum primary bed temperature is 1400°F and that a minimum temperature of about 1325°F is necessary to eliminate this discontinuity.

Growth Evaluation

Utility gas turbines are currently operating at temperatures of 1800°F to 1900°F for base load applications. Therefore, there is a possibility for improvement of the high-pressure fluidized bed system if the bed temperature is increased above 1750°F or if the temperature difference between the bed and the turbine inlet is reduced. In addition, it is thought that the boiler tube corrosion problem in the fluidized bed will be substantially less than it is in the conventional boiler so that it may be possible to increase the steam superheat temperature. Cycle calculations were made to determine the effect of increasing gas turbine temperature and/or steam superheat temperature on the plant capacity and heat rate. The results of those calculations are given in Table 1-40. This indicates a significant growth potential for the high-pressure fluidized bed boiler system.

Boiler Subsystem

A boiler pressure of 10 atm gives the best plant performance and cost for an assumed constant boiler cost. An analysis was made of the effect of pressure on boiler size and pressure vessel costs which showed that very little cost reduction could be realized by increasing the pressure level beyond 10 atm.

A four-module boiler design must be used to meet the turn-down requirements for the plant if the minimum bed temperature is 1400°F. Other design features which had to be considered are:

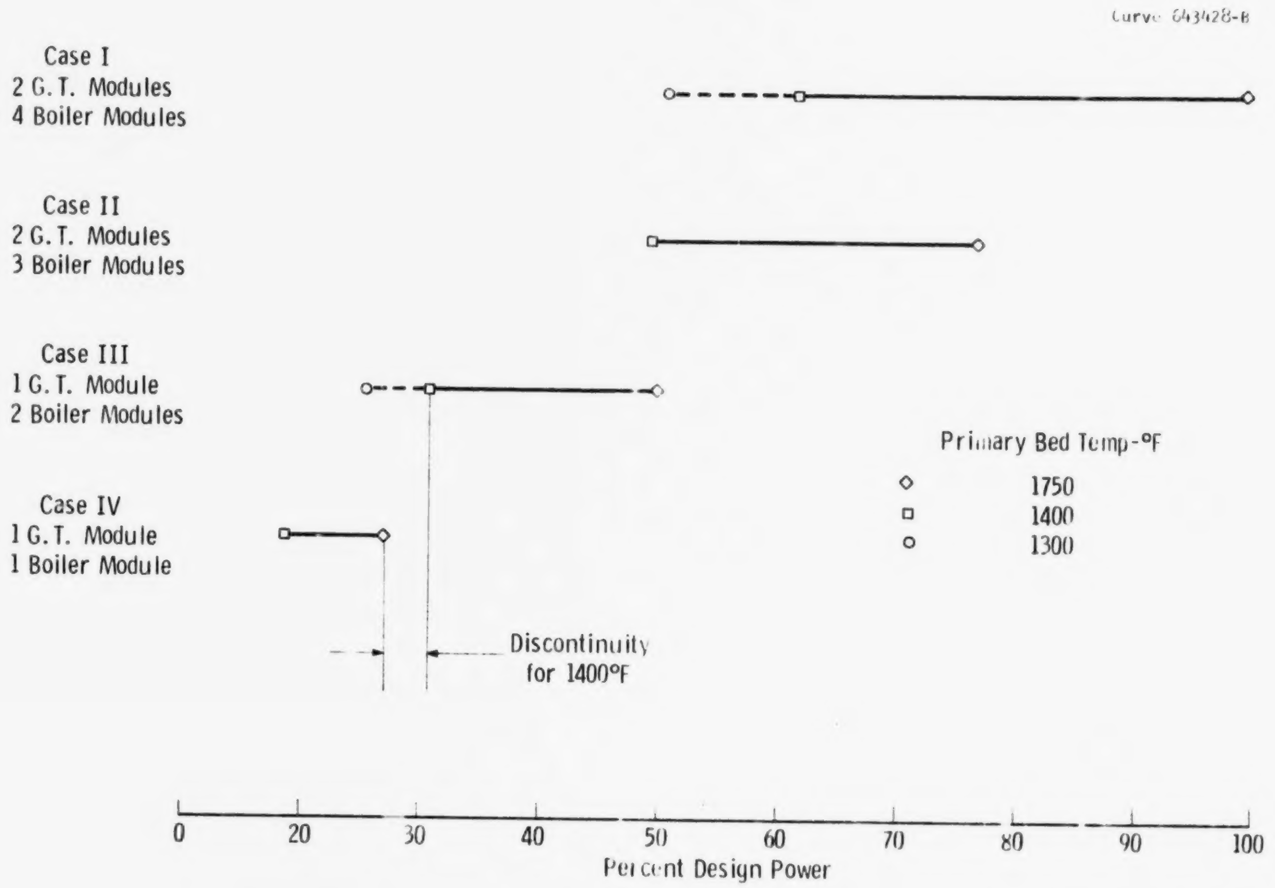


Fig. 1-103—Turn-down characteristics of combined— cycle with fluidized bed boiler operating on mode 3

Table 1-40

EFFECT OF GAS TURBINE AND STEAM SUPERHEAT TEMPERATURE ON THE HIGH-PRESSURE SYSTEM CAPACITY AND HEAT RATE

STEAM PRESS. - PSI	SUPERHEAT TEMP. °F	REHEAT TEMP. °F	GAS TURBINE TEMP. °F	GAS TURBINE FLOW %	STEAM TURBINE FLOW %	PLANT CAPACITY %	HEAT RATE %
2400	1000	1000	1600	100	100	100	100
3300	1100	1100	1600	100	95.2	104.2	96.0
3300	1100	1100	1700	100	94.0	105.4	94.8
4500	1200	1100	1600	100	92.2	106.6	93.8
4500	1200	1100	1700	100	91.0	107.8	92.7

- Arrangement of the pressure parts
- Size and shape of the fluidized bed or beds
- Arrangement of the bed or beds in the pressure vessel
- Bed depth
- Orientation of the vessel.

Since the maximum allowable bed temperature is less than the state-of-the-art gas turbine temperature, there is no need for convection heat transfer between the bed and the gas turbine. Therefore, all of the boiler heat transfer surface is located in the bed or immediately above the bed where most of the heat is absorbed from the elutriated solids rather than from the gas. In order to provide flexibility in the operation of the boiler module, each boiler function, e.g., evaporation, was allocated at least one separate bed. The level of superficial velocity used in the primary beds results in the elutriation of a significant amount of unburned carbon. Rather than recycle this material to the primary beds, a separate bed or carbon burn-up cell is used in each module. Each module, therefore, would have a minimum of four primary beds plus the carbon burn-up cell.

Alternative Concepts for Boiler Preliminary Design. Having selected a modular boiler with multiple beds in each module, choices had to be made regarding the orientation of the modules, the shape of the bed, the bed depth, and the orientation of the boiler tubes.

The vertical module with stacked bed was chosen over the horizontal bed with side-by-side beds because of its advantages in the distribution of the air to the beds, the collection of the products of combustion from the beds, and the steam circuitry arrangement. The vertical module with stacked beds also permits the use of one continuous water wall enclosure for all of the beds. The use of a horizontal module places a definite limit on the maximum bed depth for a given pressure vessel diameter.

The shape of the bed and the tube orientation are interrelated: an annular bed precludes the use of horizontal tubes in serpentine bundles; vertical tubes fit well into a circular bed, whereas horizontal tubes are possible but expensive because of the nonuniformity of tube bank element; square beds, of course, work equally well with vertical and horizontal tubes. Horizontal tubes were chosen over vertical tubes because of the possibility of poor hydrodynamics in the bed with vertical tubes and because their use permits conventional fabrication methods. This led to the choice of a rectangular bed as the preferred arrangement.

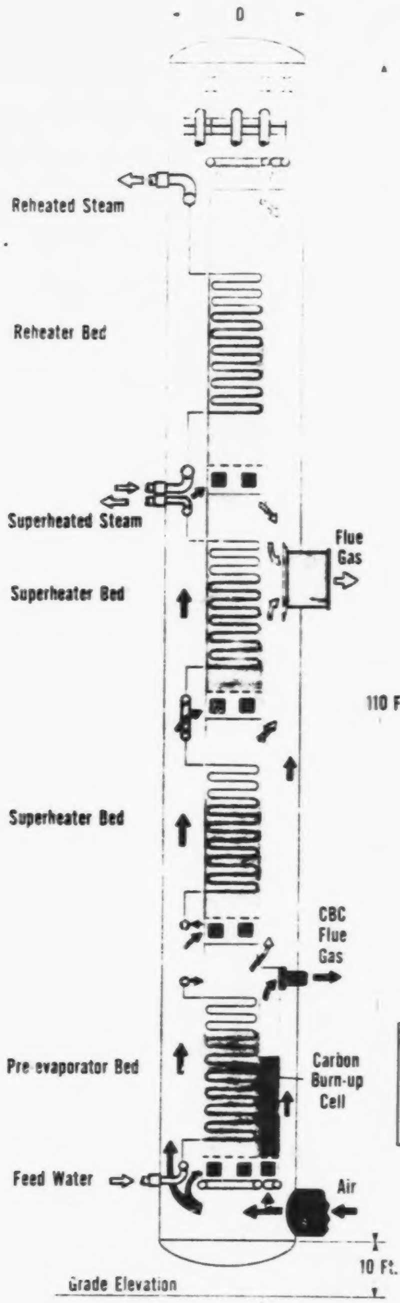
Boiler Design

Modules. An elevation drawing of the final design of the high-pressure fluidized bed boiler module is shown in Figure 1-104. Detailed drawings of the module are given in Appendix H . For a 318 MW plant module the pressure vessel is 12 feet in diameter and 110 feet high. Both of these dimensions are within their respective limits for rail transport, so the module can be shop-fabricated. The module is supported by external steel framing.

Beds. The primary beds are 5 ft x 6-1/2 ft in the plan view. By using beds of these dimensions and by using the water walls for evaporation, it was possible to get sufficient heat transfer surface for the other boiler functions in four primary beds approximately 12 ft in depth. One bed is for pre-evaporation, two are for superheating, and the fourth is for reheating.

The carbon burn-up cell is located near the bottom of the module in a water-walled enclosure adjacent to the pre-evaporator bed and has dimensions of 1-3/4 x 7 x 20 ft. There are no boiler tubes in this bed.

Boiler Tubes. Several alternative tube configurations were considered. Those selected are as follows:



PLANT SIZE	VESSEL DIAMETER, D
320 mw	12 Ft.
635 mw	17 Ft.

ELEVATION

PRESSURIZED FLUIDIZED BED STEAM GENERATOR
FOR COMBINED CYCLE PLANT
FOUR (4) REQUIRED

Fig. 1-104

FUNCTION	TUBE DIA. - IN.	SPACING - IN.	SURFACE - FT^2
Pre-evaporator	1-1/2 O.D.	3-1/2 between rows 3 in rows	13,900
Evaporator	2 in. O.D.	3-1/2 on centers in fin wall	17,200
Superheater	1-1/2 O.D.	3-1/2 between rows 3 in rows	36,400
Reheater	2 in. O.D.	3-1/2 between rows 4 in rows	16,700
			84,200

From the standpoint of fluidization, economy of space would be promoted by using the smallest diameter tubes possible; but steam flow requirements and structural problems must be considered in choosing the tube sizes. The mean metal temperatures for the boiler pressure parts were calculated, and material specifications were made for use in estimating the boiler cost.

Steam Circulation. As previously stated, this boiler is once-through circulation. Feedwater enters the module near the bottom and flows upward through the pre-evaporator tube bank. Evaporation starts in the uppermost loops of this bank and the mixture of water and steam flows into the first of four tube wall circuits. After each circuit the mixture is transported to the bottom of the next through unheated downcomers. Thus, the two-phase flow in heated tubes is always upwards, preventing phase separation. Saturated steam leaves the fourth wall circuit and flows through the two superheater beds in series and then to the high-pressure steam turbine. The steam returns to the reheat bed between the high- and low-pressure steam turbines.

Air Circuits. The combustion air is taken from the gas turbine compressor exit to the pressure vessel and is distributed to the various beds in the module by an internal manifold which is formed by the volume between the tube wall and the pressure vessel. Air dampers are provided to control the flow to the plenum chamber below each of the fluidized bed distributor plates. The carbon burn-up cell has a separate direct air supply connection to the plenum below the bed.

Products of Combustion. The products of combustion from the primary beds leave the tube wall enclosure through tube screens located near the top of the overhead volume. These tube screens are made by offsetting alternate tubes in the tube wall. The flue gases are collected in a 1-3/4 x 7 ft exhaust manifold which is formed by tube walls with one side common with the bed enclosure. The bottom 20 feet of these tube walls are the carbon burn-up cell enclosure. This is the only convection heat transfer surface in the boiler module. Near the middle of the module, an insulated pipe carries the flue gases from the waterwall passage through the pressure vessel to the first-stage particulate separator which is located outside the module.

The flue gases from the carbon burn-up cell are segregated from those from the primary beds. A separate insulated pipe leads from the carbon burn-up cell overhead out through the pressure vessel to join the primary flue gas pipe downstream of the first-stage particulate separator.

Solids Handling. Each primary bed has four injection points for pneumatically feeding coal into a tube-free space above the distributor plate. The make-up sorbent is fed in along with coal.

Each bed is equipped with one nearly vertical pipe at a point about four feet below the top of the tube bank which removes the spent sorbent from the bed for regeneration. Each bed also has one inlet pipe located about four feet above the air distributor plate for pneumatically injecting the regenerated sorbent.

The carbon burn-up cell is similar to the primary beds in that it has four injection points for the fuel, one withdrawal pipe for the spent sorbent, and one return pipe for the regenerated sorbent.

Boiler Performance

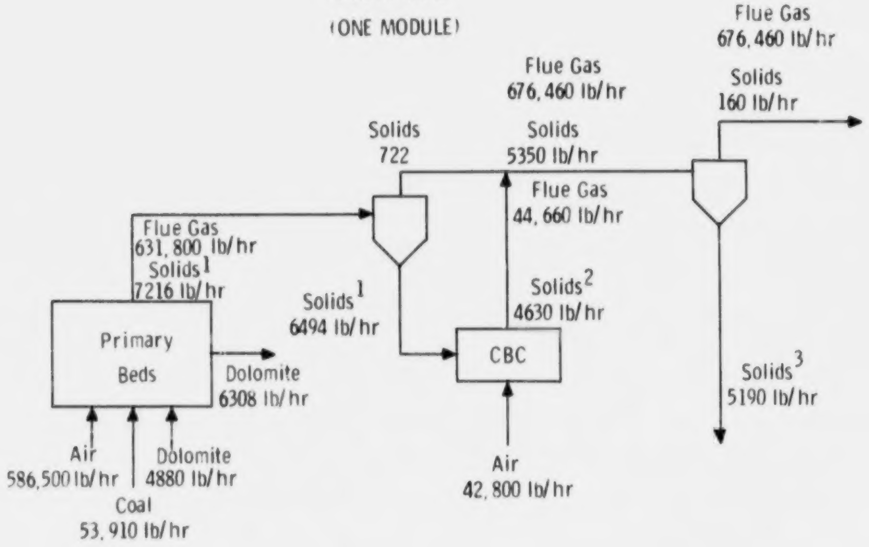
Energy and mass balances were calculated for the design load and for 70% of design load. These are shown in Figures 1-105 through 1-108.

FIGURE 1-105—MASS BALANCE

DESIGN LOAD

(ONE MODULE)

Dwg. 6168A70



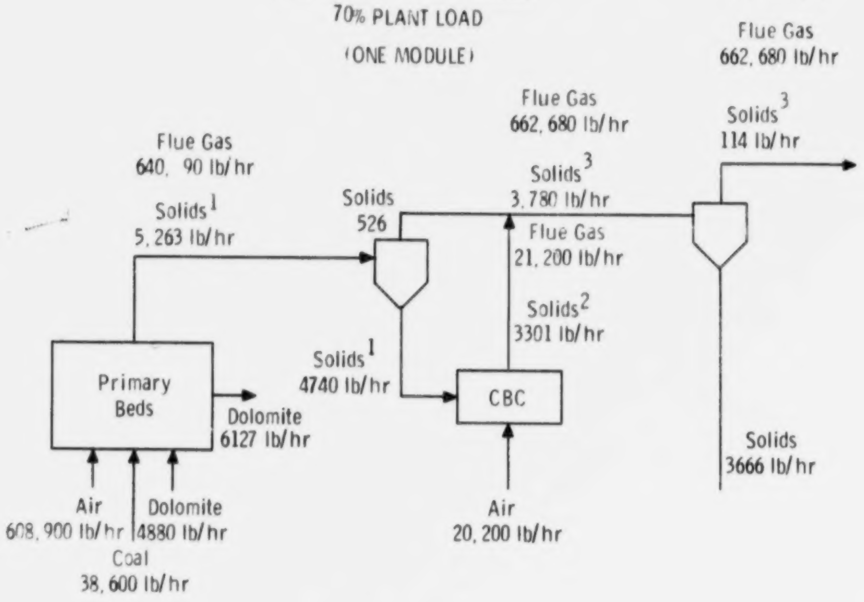
1		2		3	
Carbon	31.9%	Carbon	4.5%	Carbon	8.2
Ash	63.4%	Ash	89.0	Ash	85.8
Dolomite	4.6	Dolomite	6.5	Dolomite	6.0

FIG. 1-106—MASS BALANCE

70% PLANT LOAD

(ONE MODULE)

Dwg. [illegible]



1		2		3	
Carbon	31.2%	Carbon	4.3%	Carbon	7.0%
Ash	62.1%	Ash	86.4%	Ash	83.9%
Dolomite	6.7%	Dolomite	9.3%	Dolomite	9.1%

FIG. 1-107-ENERGY BALANCE

DESIGN LOAD

(ONE MODULE)

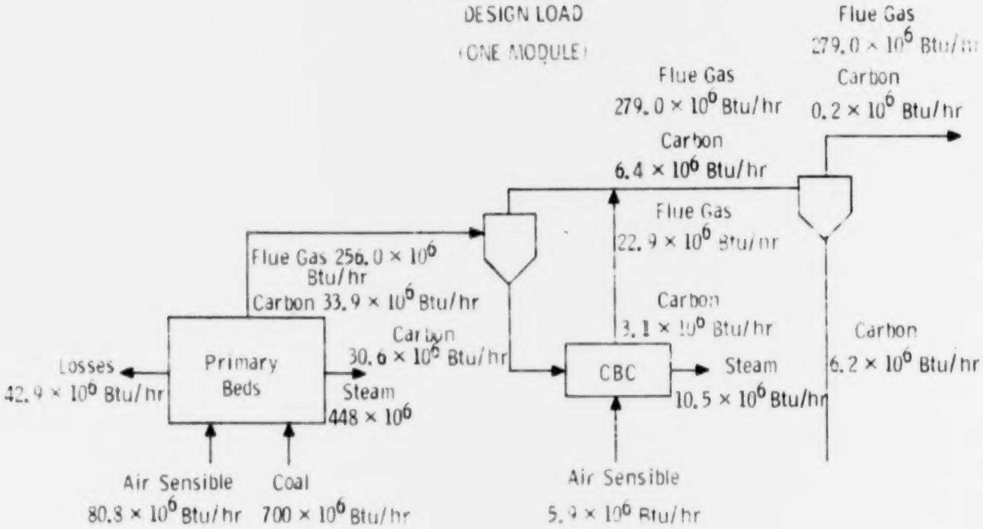
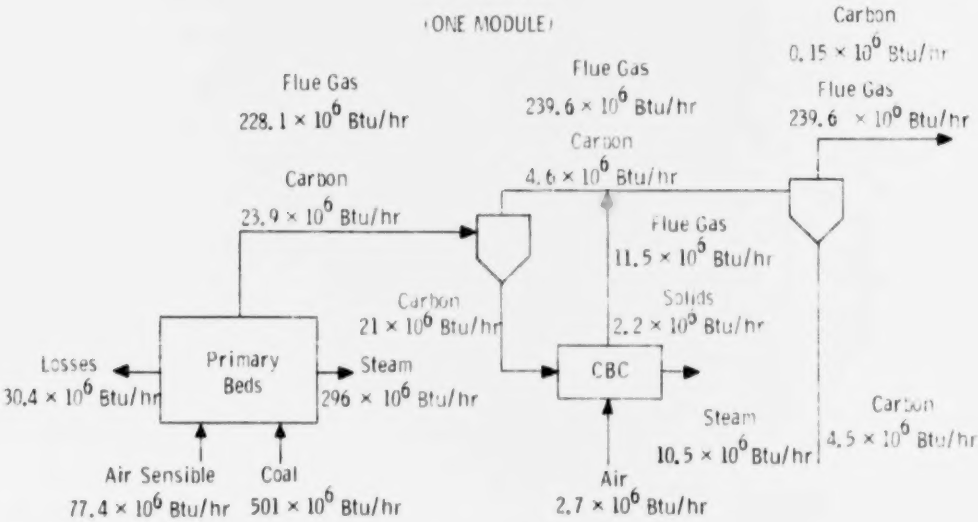


FIGURE 1-108-ENERGY BALANCE

70% Plant Load

(ONE MODULE)



Steam generator efficiency calculations are misleading in a combined cycle plant if the gas temperature leaving the steam generator is used to calculate the stack gas losses. The stack gas temperature should be used to calculate a combination boiler and gas turbine combustor efficiency. The calculated and estimated losses for the high-pressure fluidized bed boiler on this basis are as follows:

COMPONENT	LOSS - %
Dry gas loss @ 275°F	3.88
Loss due to hydrogen and moisture in coal	4.14
Loss due to moisture in air @ 275°F	0.08
Radiation loss	0.15
Incomplete combustion	1.51
Solids sensible heat	0.11
Unaccounted for and manufacturer's margin	1.50
	11.37

This gives a combustion efficiency of 88.6%.

Regeneration/Sulfur Recovery System

The regenerator system selected for the high-pressure boiler is based on the "reduction-steam CO_2 oxidation" reaction scheme. Process options and thermodynamics were discussed above, and Figure 1-109 is a flow diagram of the selected process. Table 1-41 gives material balance information, and Figure 1-110 is a schematic plant layout showing boiler-regenerator interconnections.

Spent stone discharges from each boiler bed and the carbon burn-up cells to a hold vessel; one hold vessel is provided for each boiler module. Four lines pneumatically convey spent stone from each hold vessel to the first-stage regenerator reactor -- the $CaSO_4$ reducer vessel. Air serves as the carrier gas. The reactor, a fluid bed reactor, operates at 1500°F and 8 atm. The gas producer vessel supplies reducing gas for $CaSO_4$ reduction by partially combusting coal with air and steam. After particulate removal, the effluent gas from the

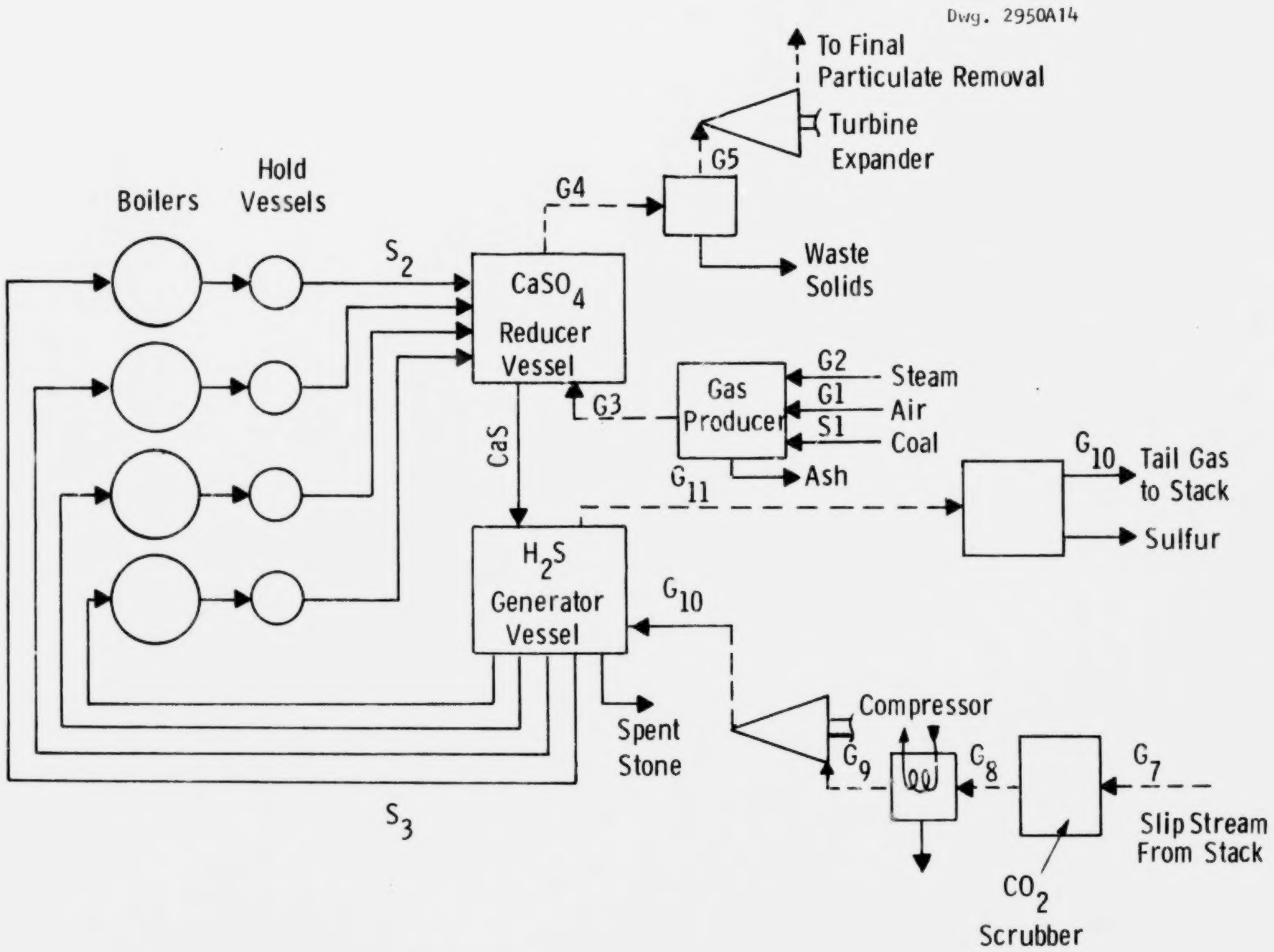


Fig. 1-109—Pressurized regeneration System—Flow Diagram

TABLE 1-41

600 MW HIGH-PRESSURE REGENERATOR MATERIAL BALANCE

STREAM	T (^oF)	P (PSIA)	LB/HR	LB MOLE / HR	MOLE %						
					H_2	H_2O	CO	CO_2	N_2	O_2	H_2S
G1	640	150	183,000	6,300	--	--	--	--	79.1	20.9	--
G2	640	150	9,000	500	--	100	--	--	--	--	--
G3	1500	135	221,000	8,400	8.0	8.4	16.3	7.4	59.4	--	0.5
G4	1500	120	182,000	6,800	1.0	19.8	0.3	5.5	73.4	--	--
G5	1500	116	182,000	6,800	1.0	19.8	0.3	5.5	73.4	--	--
G6		∿18	182,000	6,800	1.0	19.8	0.3	5.5	73.4	--	--
G7(a)	230	135	595,000	20,000	--	8.6	0.2	16.0	73.4	1.7	--
G8(a)	220	19.7	266,000	10,000	--	68.8	--	30.9	0.2	--	--
G9(a)	212	19.0	179,000	6,500	--	63.7	--	36.3	--	--	--
G10(a)	>212	180	179,000	6,500	--	63.7	--	36.3	--	--	--
G11	1100	165	117,000	4,900	--	73.9	--	16.1	--	--	10.0
S1 (coal)	Amb	135	33,500								
S2 (spent dolomite)	1500	135	400,000	∿16% sulfated solids							
S3 (regenerated dolomite)	1100	180	450,000	∿1% sulfated compositions							

(a) Flow rate for CO_2 scrubber design basis.

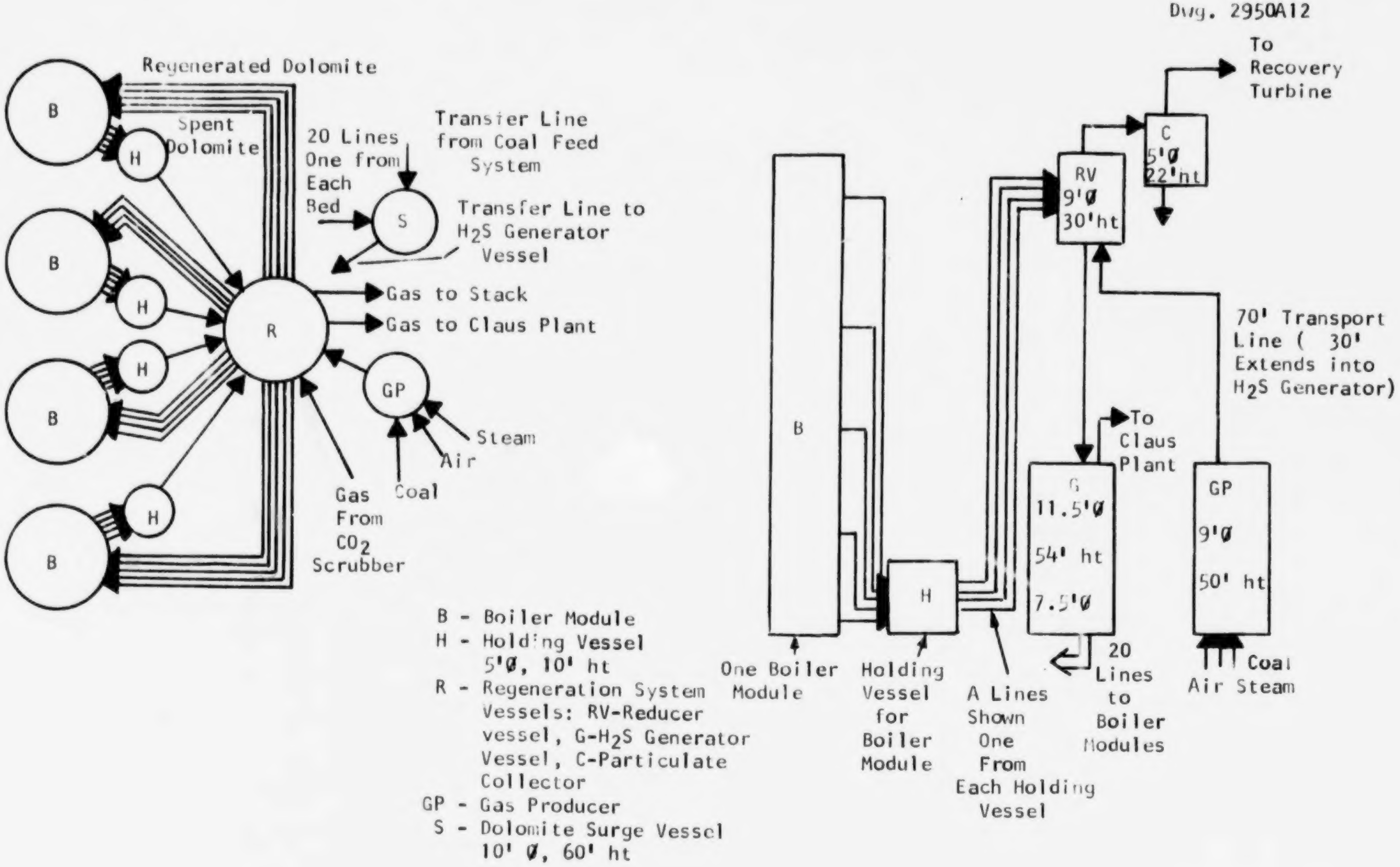

Dwg. 2950A12
Regenerated Dolomite
Spent Dolomite
B
H
20 Lines One from Each Bed
Transfer Line from Coal Feed System
S
Transfer Line to H2S Generator Vessel
R
Gas to Stack
Gas to Claus Plant
GP
Steam
Air
Coal
Gas From CO2 Scrubber
B - Boiler Module
H - Holding Vessel 5'0, 10' ht
R - Regeneration System Vessels: RV-Reducer vessel, G-H2S Generator Vessel, C-Particulate Collector
GP - Gas Producer
S - Dolomite Surge Vessel 10' 0, 60' ht
One Boiler Module
Holding Vessel for Boiler Module
A Lines Shown One From Each Holding Vessel
RV 9'0 30'ht
C 5'0 22'ht
To Recovery Turbine
70' Transport Line (30' Extends into H2S Generator)
To Claus Plant
G 11.5'0 54' ht 7.5'0
20 Lines to Boiler Modules
GP 9'0 50' ht
Coal
Air Steam

Fig. 1-110—Pressurized regenerator System: Plant Layout

$CaSO_4$ reducer vessel is expanded through a turbine expander which drives a compressor. The effluent gas passes through a final particulate removal step before being discharged to the power plant flue gas.

Reduced solids, CaS, flow through a standpipe leg into the second-stage reactor, the H_2S generator vessel. This fluidized bed operates at 1100°F and 12 atm. The standpipe serves as a pressure seal between the reducer and the generator vessels. CO_2 and steam convert CaS to $CaCO_3$ for recycle to the boiler.

Regenerated solid is conveyed back to each boiler bed through 20 discharge lines from the H_2S generator — one line per boiler bed and CBC. Feed gas to the second reactor is provided by scrubbing a CO_2-rich stream of boiler flue gas in a regenerable hot carbonate or monoethanol amine (MEA) scrubber system. The CO_2 compressor is driven by the turbine expander on the reducer vessel outlet. H_2S-rich gas is sent to a conventional Claus plant for conversion to elemental sulfur.

Overall regeneration performance, particularly with respect to sulfur removal efficiency, is impossible to predict precisely. Conventional Claus plants recover 94% to 96% of the sulfur fed to them. With added tail gas clean-up this can be improved to 99+%. Losses from the regenerator should amount to approximately 2% to 5% of the total sulfur (assuming gas leaving the reducer vessel has the equilibrium SO_2 concentration). Therefore, the overall sulfur recovery efficiency should be around 90%.

The cost estimate for the regeneration system for a 635 MW plant is \$8.90/kw. The Claus plant system cost is \$3.78/kw. Claus plant tail gas clean-up will add ~ \$3/kw.

The cost estimates for the various components of this regeneration system were obtained from the following sources. Detailed design information is given in Appendix I.

• Solids feed system	Petrocarb
• Claus plant	Ford, Bacon and Davis Texas

• CO_2 scrubber	Benfield Corp.
• Process vessels	Westinghouse Heat Transfer Div.

Auxiliary Equipment

Coal Handling. The coal handling system includes receiving, storage, drying, crushing, and conveying facilities (See Figure 1-111). Coal (1-1/2 x 0 in) arrives at the plant by unit train twice a week and is unloaded into the 400-ton receiving hopper in about five hours. A 12,000-ton silo acts as a buffer between the batch-type delivery operation and the continuous feeding operation. It also provides live storage for short-term delays in coal deliveries. Dead storage for long-term delays is provided by a 50,000-ton open pile. An open pit for reclaiming coal from dead storage is provided over the conveyor between the receiving hopper and the silo.

Coal from the silo, to which make-up dolomite is added, is dried to <3% moisture in a fluid bed coal-fired dryer. It is then crushed to 1/4 x 0 in. in a reversible hammermill. The size distribution of the crushed coal is given in Appendix H. It is then conveyed to a 150-ton surge-bin from which dual feeders and conveyors carry it to the interface with the coal pressurizing and feeding system.

Make-up Sorbent Handling. The dolomite sorbent is purchased crushed to size (1/4 x 0 in) and is delivered to a 100-ton receiving hopper in covered cars. A 2000-ton silo is provided with sufficient capacity for both live and dead storage. The crushed sorbent is fed onto the coal conveyor just ahead of the coal dryer.

Coal Pressurizing and Feeding. The coal pressurizing and feeding system includes feeder surge bins, lock hoppers, fuel injectors, and pneumatic transport lines (see Figure 1-112). The function of the surge bin is to receive the coal from the coal handling system and dampen fluctuations in flow rate. The automatically controlled lock hopper pressurizes the coal from one atm to the pressure required for feeding the coal to the bed. A comprehensive study of pressurizing techniques was

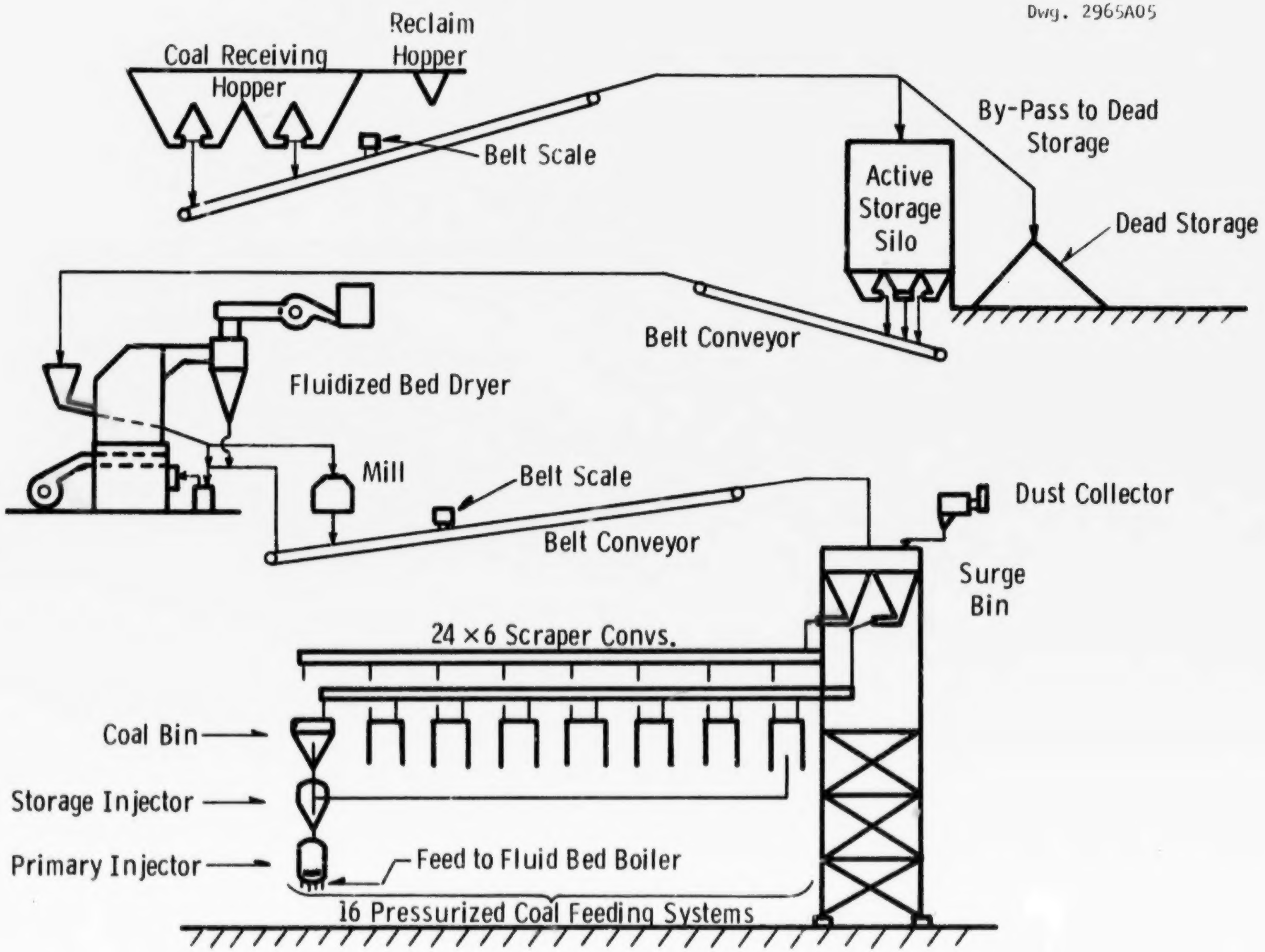


Fig. 1-111—Coal handling and feed system schematic

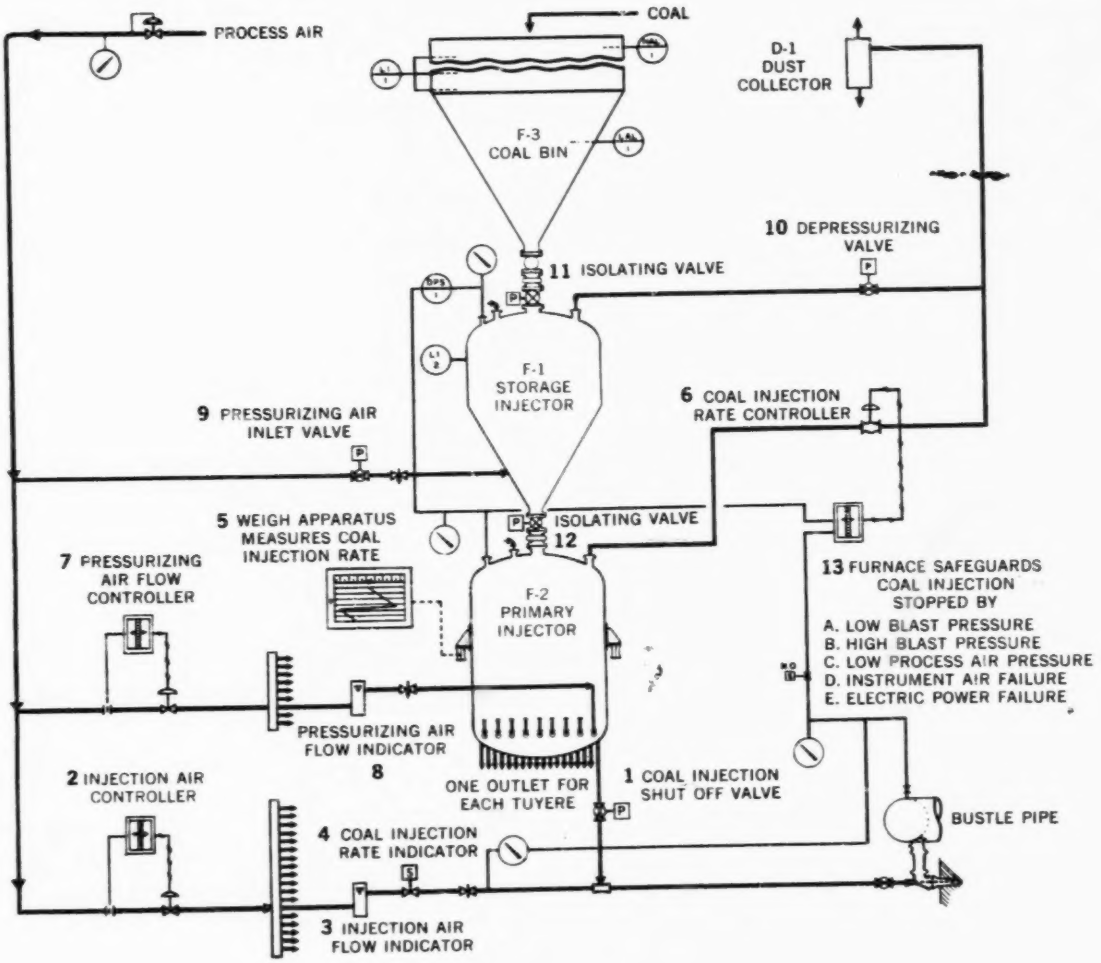


FIGURE 1-112 PETROCARB PRESSURIZED COAL FEEDING SYSTEM

carried out (see Appendix H) before the lock hopper was chosen. Next comes a fluidized bed fuel distributor which meters the coal-sorbent mixture to pneumatic transport lines which convey it to the injection tuyères.

There is a surge bin, lock hopper, distributor combination for each primary bed and a feeder outlet for each of the four injection tuyères in a bed. The pneumatic transport from the feeders to the tuyère is controlled by the amount of transport air. Compressed air requirements for the coal feed system for each of the boiler modules are estimated to be 8400 SCFM at 200 psig.

Particulate Removal. The particulate removal system in the combined-cycle plant with fluidized bed boilers has three functions: to recover most of the unburned fuel in the solids elutriated from the primary beds for recycle to the carbon burn-up cell; to reduce the concentration of particulates in the gas stream going to the gas turbine below the level where turbine blade erosion is a problem; and to control the emission of particulates to the atmosphere.

Figure 1-113 shows the particulate collection flow sheet for the high-pressure boiler modules. The anticipated requirement for erosion-free gas turbine operation is less severe than that projected for atmospheric emissions.

It is thought that an optimum system will be one where the second-stage separator would meet the turbine requirements but not the emission requirements; a third collector in the flue gas duct would meet the emission requirements.

Therefore, for this preliminary design, collection efficiencies were specified for the first-stage (90%) and for the second-stage (97%) collector to meet only the requirements for erosion control.

The first-stage collector is a conventional cyclone separator. Four cyclones in parallel are used for each boiler module, and these are enclosed in a single pressure vessel. The design point pressure drop for this collector is 0.7 psi.

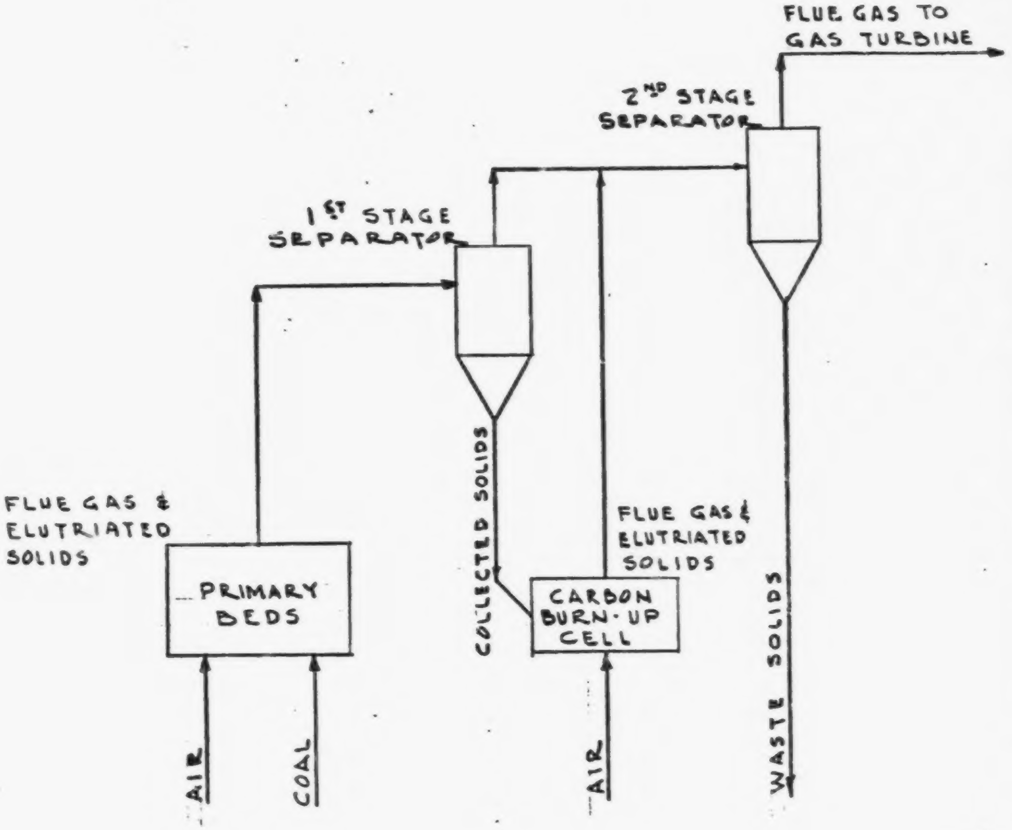


FIG. 1-113 SCHEMATIC OF PARTICULATE COLLECTION SYSTEM

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SCALE: = 1' - 0"

The second-stage collector is a tornado separator of the type shown in Figure 1-114 wherein the dirty gas is split into primary and secondary streams. All particles larger than five microns are removed in this collector. Two of these collectors are used in parallel for each boiler module. Each unit has its own pressure vessel. The arrangement of the boiler modules with the first- and second-stage particulate separators is shown in Figure 1-115.

Power Generation Equipment

Gas Turbine. Two Westinghouse W251A gas turbines match the requirements for the 318 MW combined-cycle plant. This gives two gas turbine modules with two boiler modules each.

The high-pressure fluidized bed boiler is also the combustor for the gas turbine, so provision must be made in the gas turbine design for removing the air from the compressor discharge and returning the hot gases to the turbine inlet. No industrial and utility gas turbines manufactured in the United States have external combustors; they all use an integral combustor arrangement similar to that used in aviation gas turbines. Gas turbines which are designed for use with external recuperators can be adapted to external combustors with minor modifications. Only one commercially available gas turbine is provided with a recuperator — the General Electric Series 5000 unit.

In this study of combined cycle plants using fluidized bed boilers, two Westinghouse gas turbine models have been specified. In the preliminary design of a high-pressure boiler for a 318 MW plant made by Foster Wheeler, two Westinghouse W251A gas turbines are used. In the 635MW plant for which plant layout and cost estimates were made by United Engineers and Constructors, two Westinghouse W501A gas turbines are used. In both cases, each gas turbine module has two boiler modules. The Westinghouse Small Steam and Gas Turbine Division made a layout of a modified W501A gas turbine for use with a high-pressure fluidized bed boiler and estimated the cost of manufacturing and developing this modified gas turbine design. It was assumed that the W501 design and cost information would be applicable to the W251 unit.

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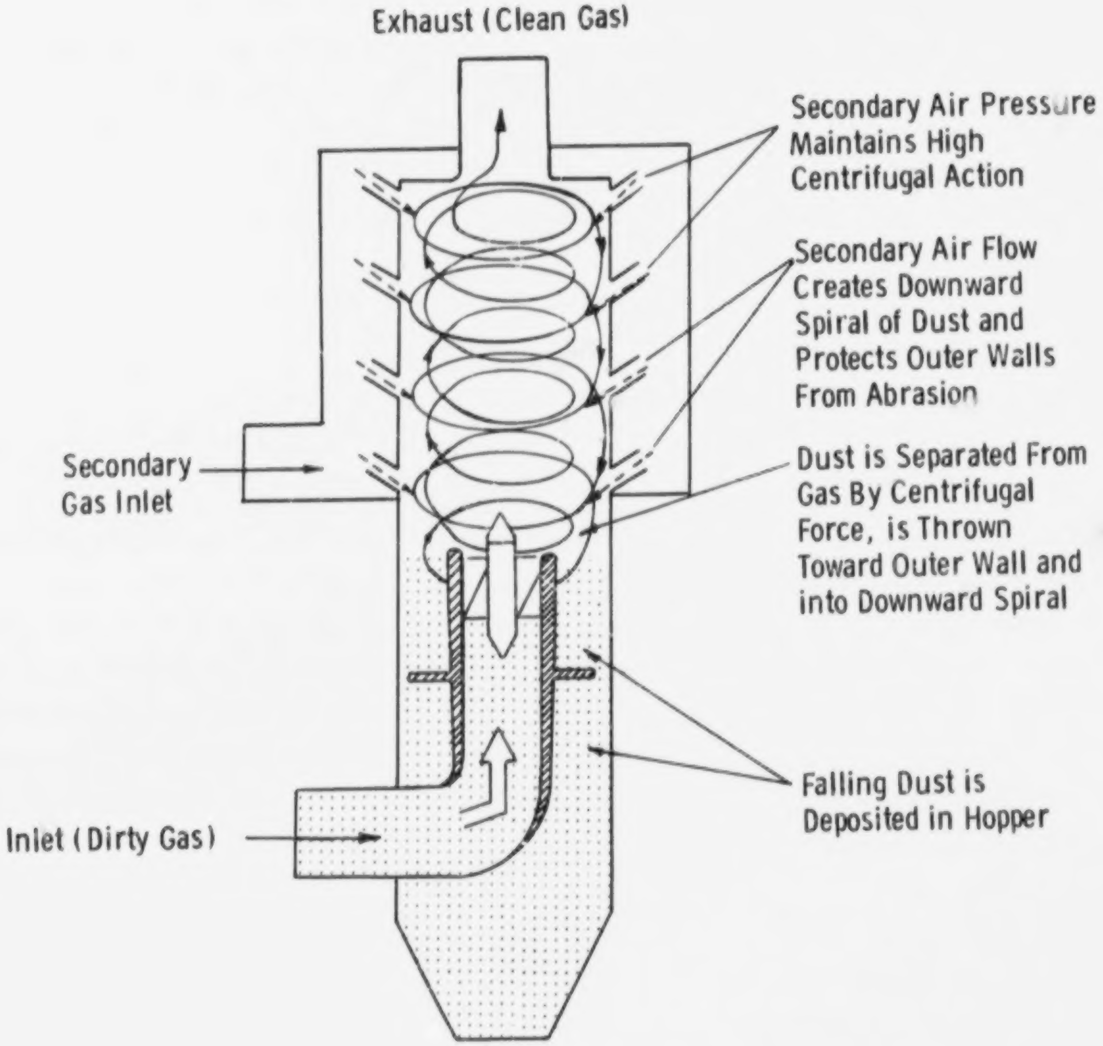
Exhaust (Clean Gas)
Secondary Air Pressure Maintains High Centrifugal Action
Secondary Air Flow Creates Downward Spiral of Dust and Protects Outer Walls From Abrasion
Secondary Gas Inlet
Dust is Separated From Gas By Centrifugal Force, is Thrown Toward Outer Wall and into Downward Spiral
Falling Dust is Deposited in Hopper
Inlet (Dirty Gas)

Fig. 1-114— Operation of Aerodyne particulate separator

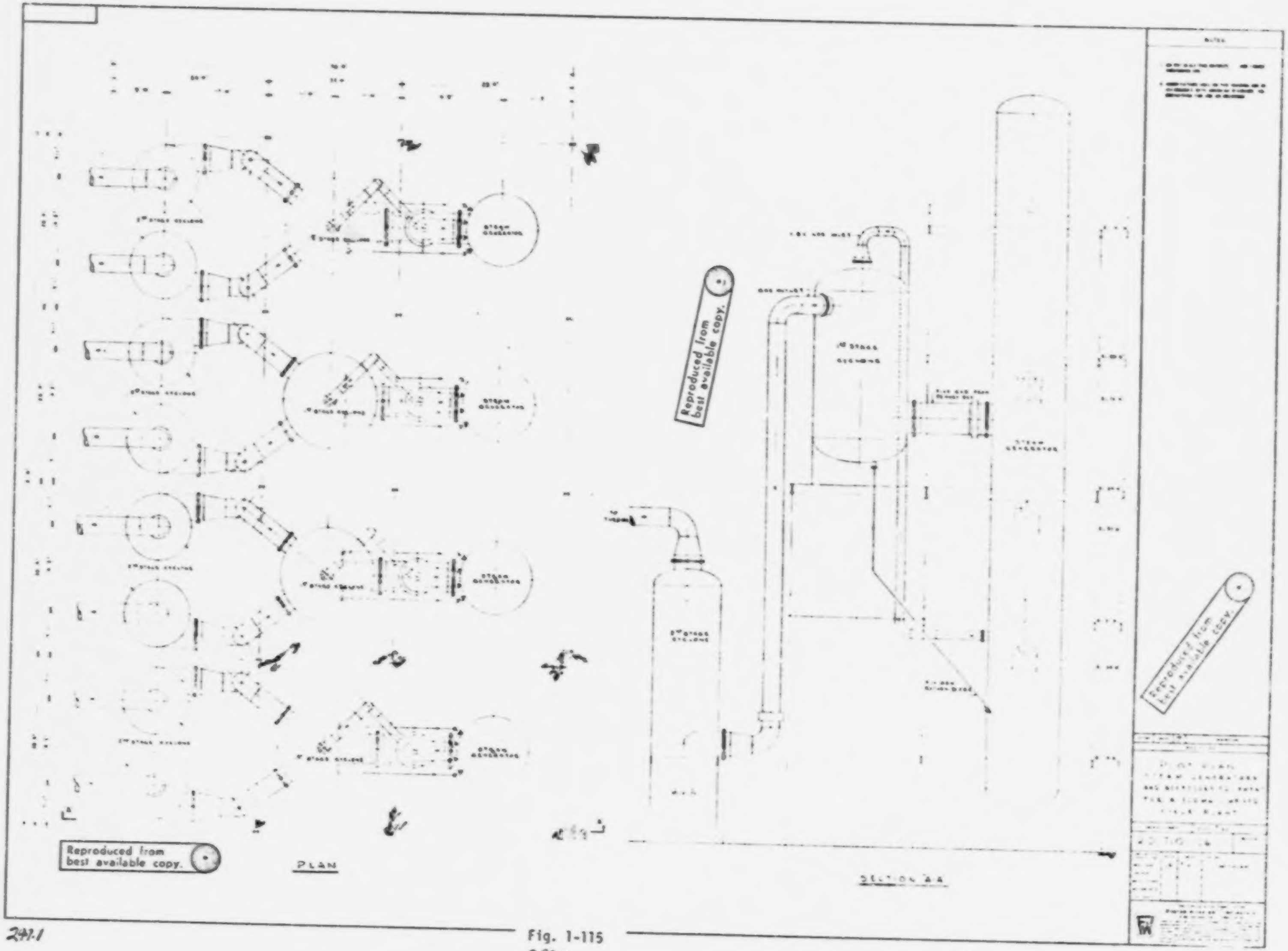

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Fig. 1-115

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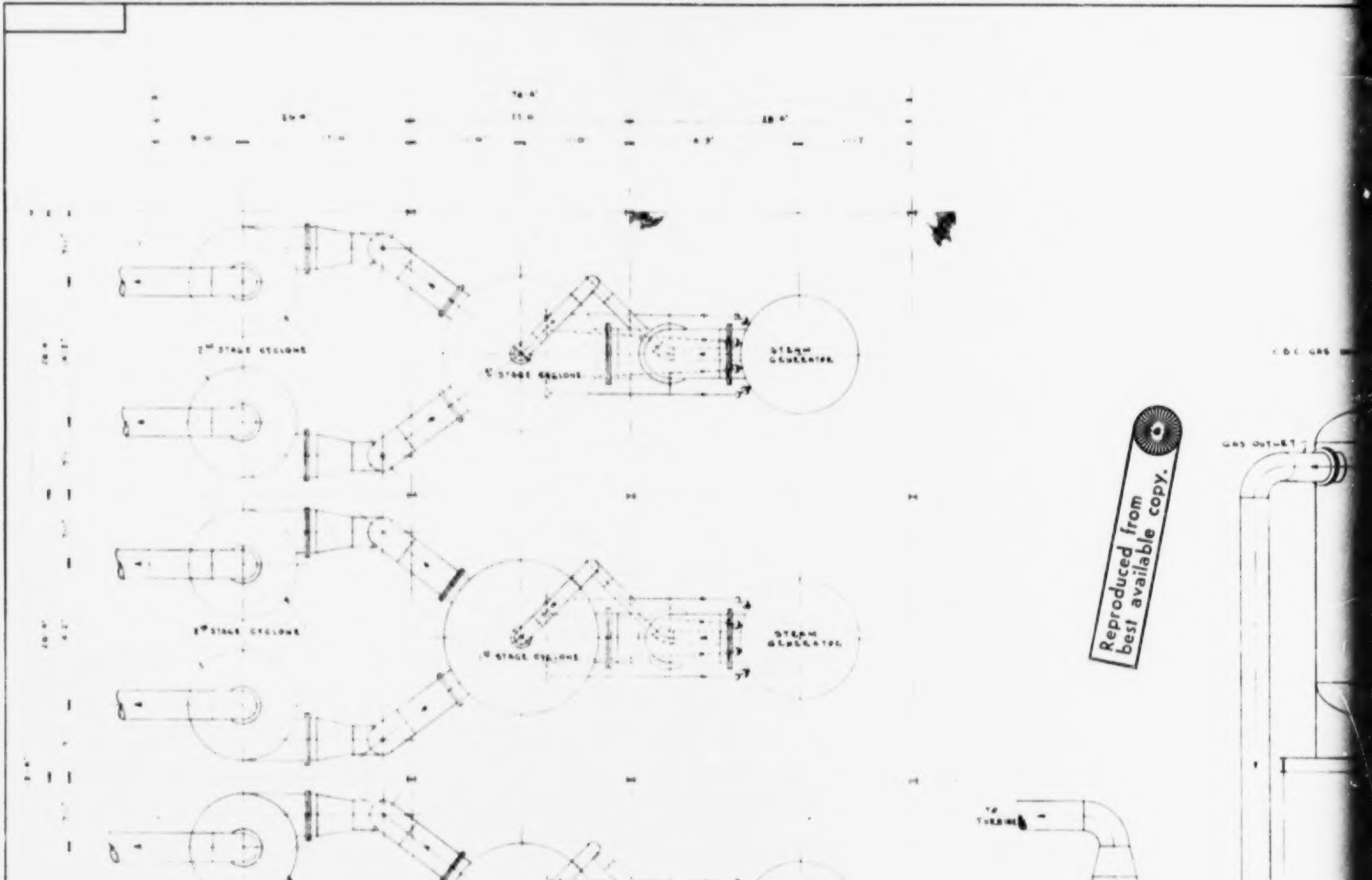

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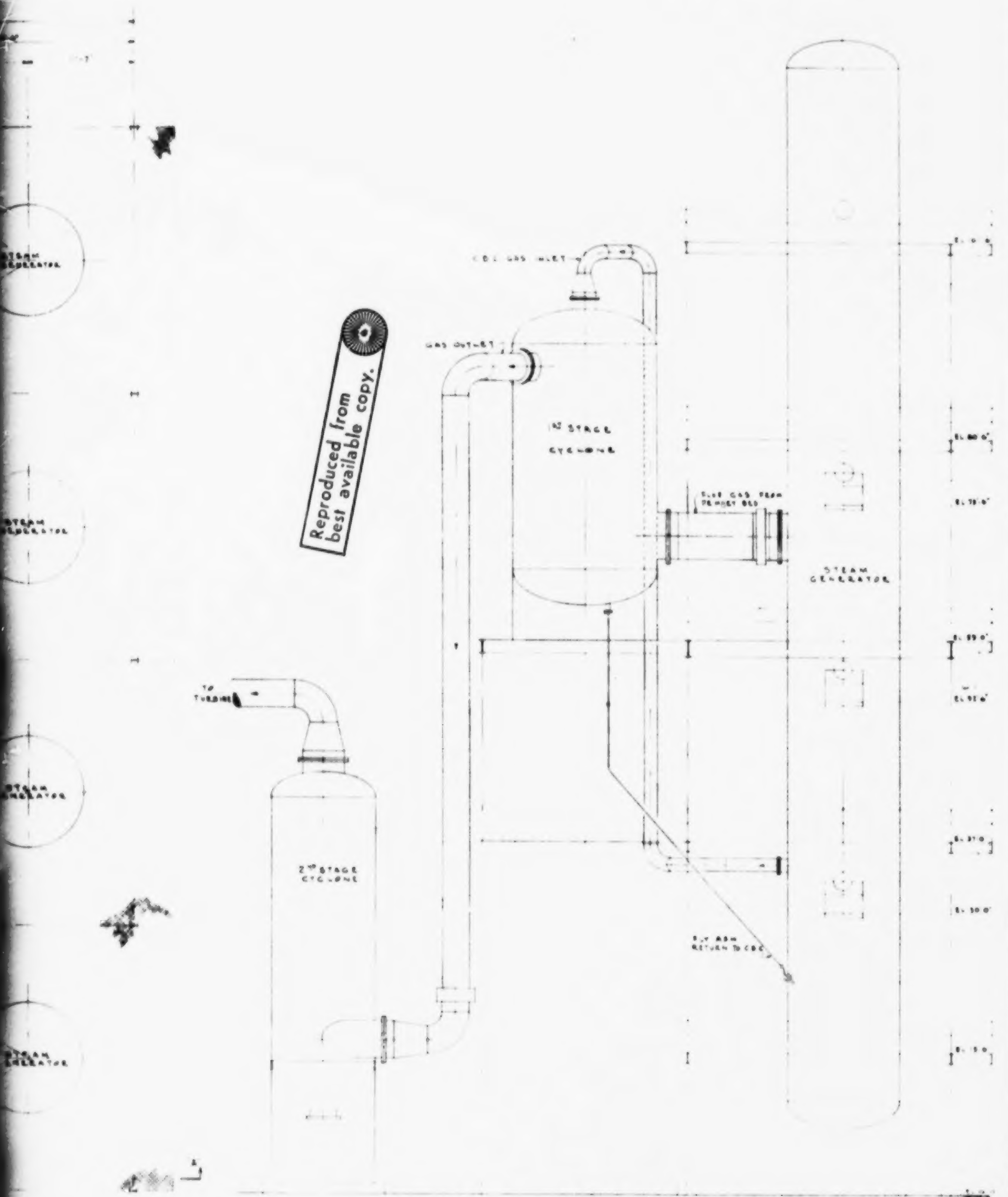
Fig. 1-115



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NOTES

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LETTER	DATE	BY	DESCRIPTION

REVISIONS

PLOT PLAN
STEAM GENERATORS
AND ACCESSORY EQUIPMENT
FOR A 300 MW COMBINED
CYCLE PLANT

DRAWING NUMBER
2D-710-106

DWG. MADE & APPROVED BY	INITIAL	DATE	CONT. NO.
DRAFTSMAN	JR	9/9/71	ORDER NUMBERS
CHECKED			
PROJECT ENGINEER			
SECTION HEAD			
DEPT. SUPV.			

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Fig. 1-115
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299.3

In the late 1950s Westinghouse developed a gas turbine for operation on blast furnace gas. One unit was built for U. S. Steel and installed at the Chicago South Works. This gas turbine had an external can-type combustor located vertically below the unit which supplied hot gas to the turbine at 1350°F. Figure 1-116 shows the internal arrangement for removing the compressor air and returning the products of combustion to the turbine inlet. There are two air connections which are isolated from the hot gas manifold; the hot gas manifold has a single connection. This unit has operated for 16,000 hours on blast furnace gas and for 12,000 hours on natural gas.

In the early 1960s, FIAT [16], an Italian licensee of Westinghouse, developed a 40 MW compound cycle gas turbine which uses external combustors (see Figure 1-117). In 1968 [104], three of these units were in operation and four were on order. Westinghouse assisted FIAT in the development of these combustors, which were patterned after the blast furnace combustor design. The early units were 1350/ 1350°F, and the later units were 1450/ 1450°F.

The low-pressure unit has two air connections and a single hot gas connection to the internal manifold, the same as the blast furnace unit. The high-pressure unit has a single concentric connection for both the air and the hot gas.

By mid 1968, each of the three units installed had accumulated about 10,000 hours of operation on residual and distillate fuels.

Brown Boveri [104], Siemens [81], and Kraftwerk Union AG [82] also build gas turbines with external combustors. The Brown Boveri units are compound cycles with single combustors and operate mostly at 1150°F. The Siemens unit, which is being used in a Velox boiler application, has a turbine inlet temperature of about 1500°F. Its configuration is not known. The Kraftwerk-Union turbine operates at 1475°F and has two external combustors with concentric air and gas connections (see Figure 1-118).

While the above-mentioned gas turbines with single or double connections have operated satisfactorily, the turbine inlet temperatures used are appreciably lower than those anticipated for this application.

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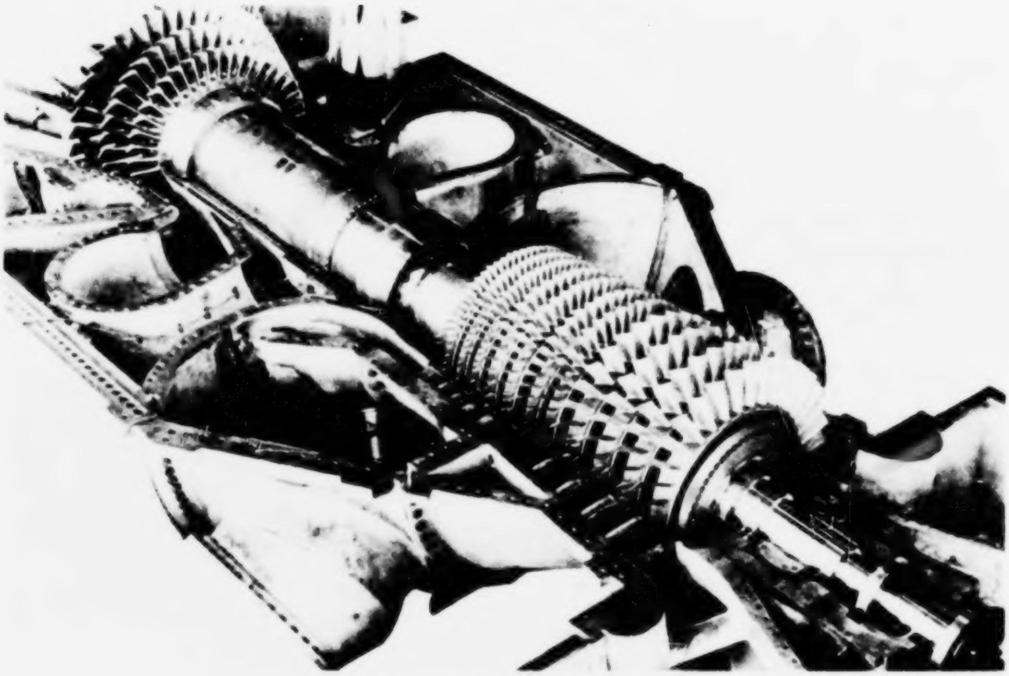
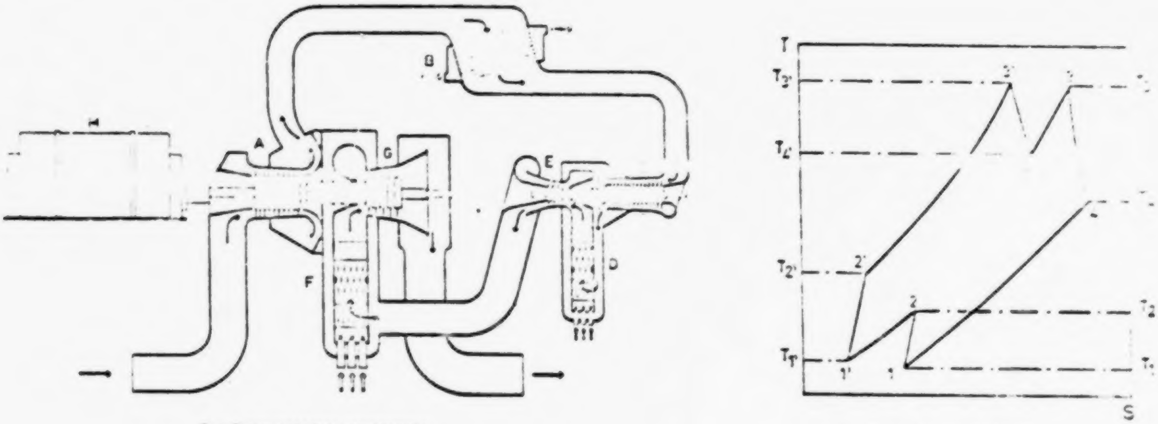


Fig. 1-116



A - Low-pressure compressor
B - Air intercooling
C - High-pressure compressor
D - High-pressure combustor
E - High-pressure turbine
- Low-pressure combustor
G - Low-pressure turbine
M - Load

1 - 2 - Air compression
2 - 1' - Air cooling
1' - 2' - Air recompression
2' - 3' - Heating of air
3' - 4' - Expansion of hot gases
4' - 3 - Reheating of gases
3 - 4 - Expansion of hot gases
4 - 1 - Exhaust of gases

FIGURE 1-117 TG3000 COMPOUND-CYCLE GAS TURBINE SCHEMATIC CYCLE ARRANGEMENT AND OPERATING DIAGRAM

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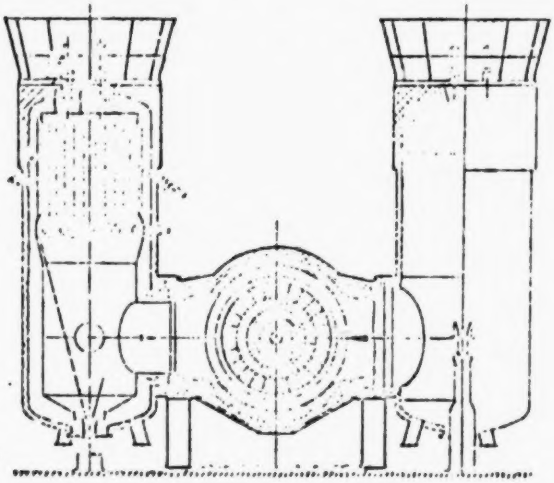

Fig. 1-118 — Gas turbine V95. Section through combustion chamber

For gas turbine inlet temperatures of 1600°F and greater, distortion problems are anticipated with internal hot gas manifolds because of non-uniform temperatures. Therefore, the Westinghouse Small Steam and Gas Turbine Division designed a transition system without an internal hot gas manifold (see Figures 1-119, 1-120, and 1-121).

The system has 16 concentric air and hot gas pipes which lead from the unit housing between the compressor exit and the turbine inlet to two vertical external manifolds. Separate connections for the air and the hot gas are provided at these external manifolds. A crossover pipe between the two manifolds is provided for the hot gas. The hot gas pipes terminate at the turbine inlet in 16 sectors which completely fill the turbine inlet annulus, thus eliminating the need for an internal hot gas manifold.

It has not been concluded that an internal manifold with one or more connections is not feasible for use at turbine inlet temperature over 1600°F. Such a transition system could be less complex and possibly less expensive than the one shown in Figures 1-119 and 1-120. It is thought, however, that a substantial amount of development effort would be required to obtain an adequate internal manifold design for this temperature level, whereas the external manifold design presented herein is considered to be more nearly a state-of-the-art design.

The May, 1970 cost of a Westinghouse W501 AA gas turbine is $4,000,000. It is estimated that the manufacturing cost of a unit modified for use with the high-pressure fluidized bed boiler will be the same. The development of the modified gas turbine design is expected to take two to three years and cost about $2,000,000.

Estimated costs for the air and hot gas transition system with the external manifolds is as follows:

Concentric piping for air and hot gases	$ 48,500
Concentric air and hot gas manifolds	31,600

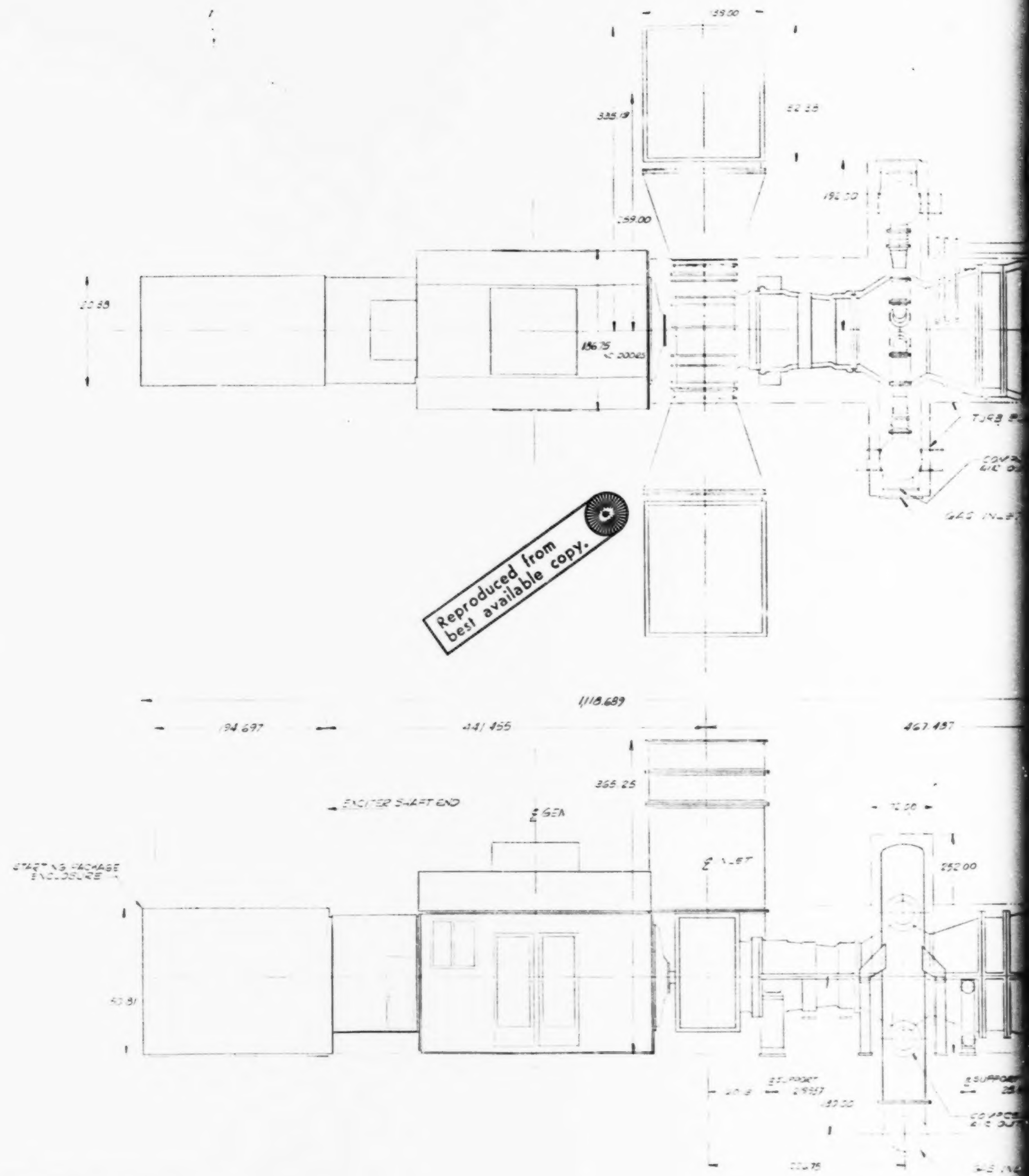

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Fig. 1-119—Plan and elevation of W501 gas turbine modified for external combustion

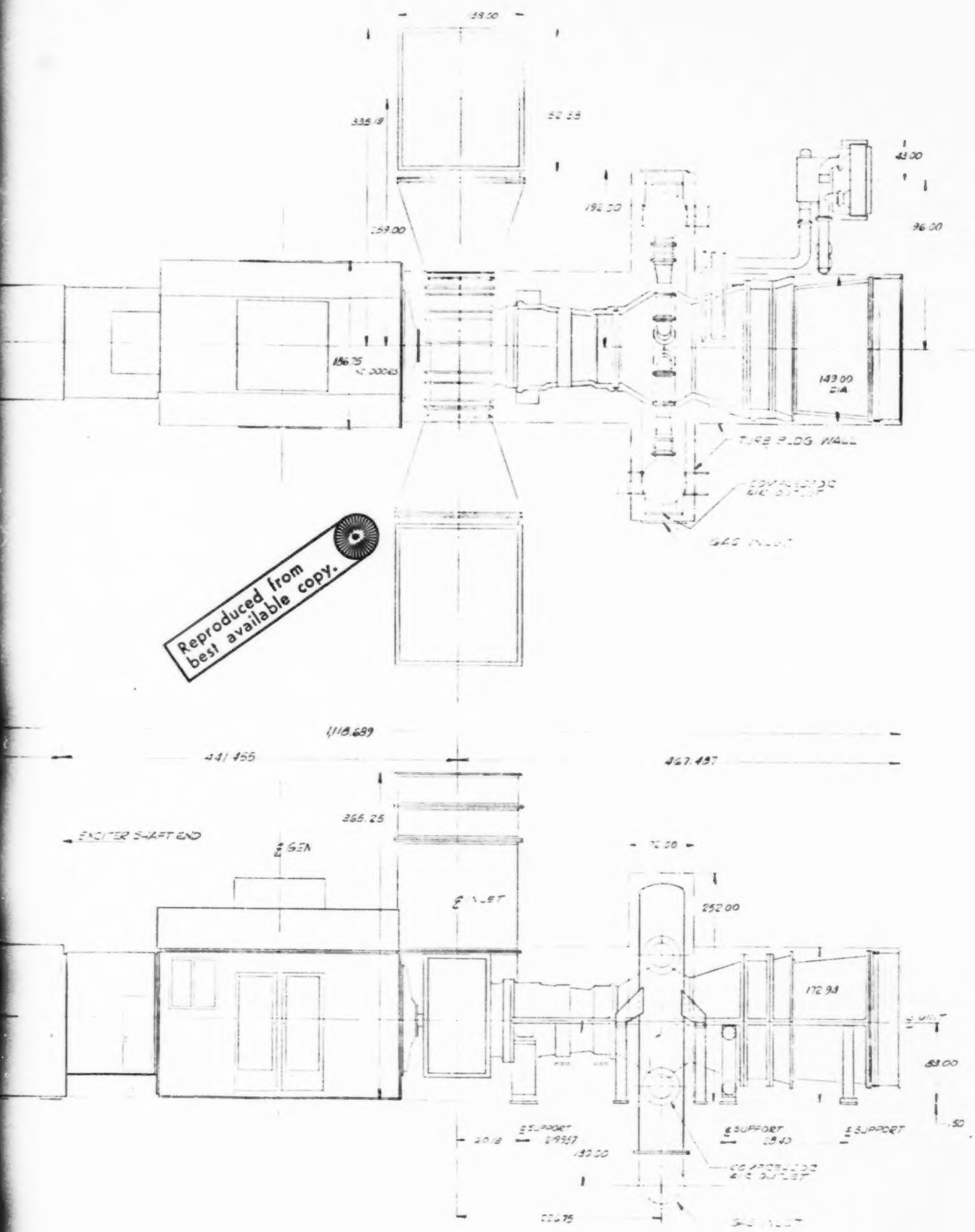

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Fig. 1-119—Plan and elevation of W501 gas turbine modified for external combustion

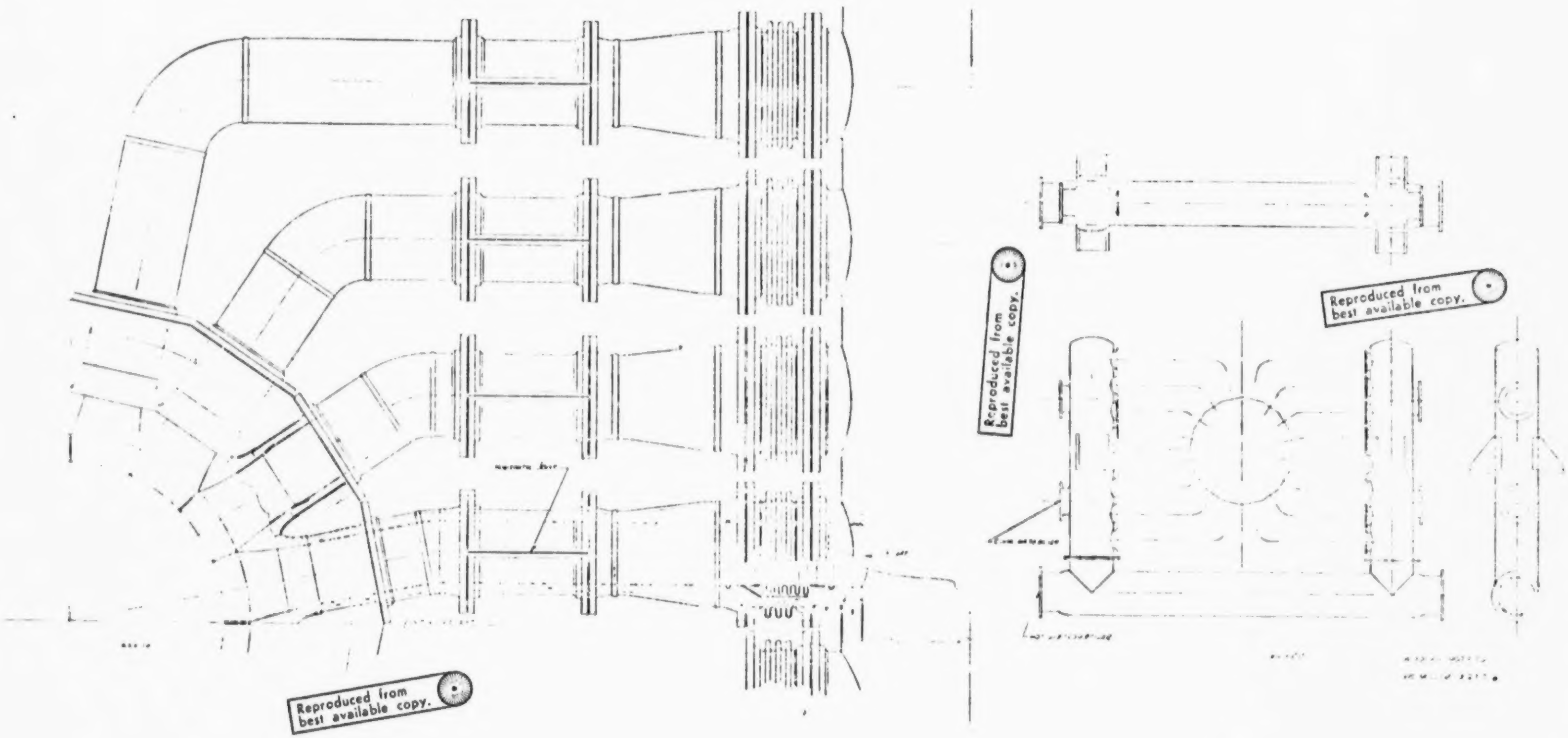

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Fig. 1-120—Detail of piping arrangement for modified W501 gas turbine

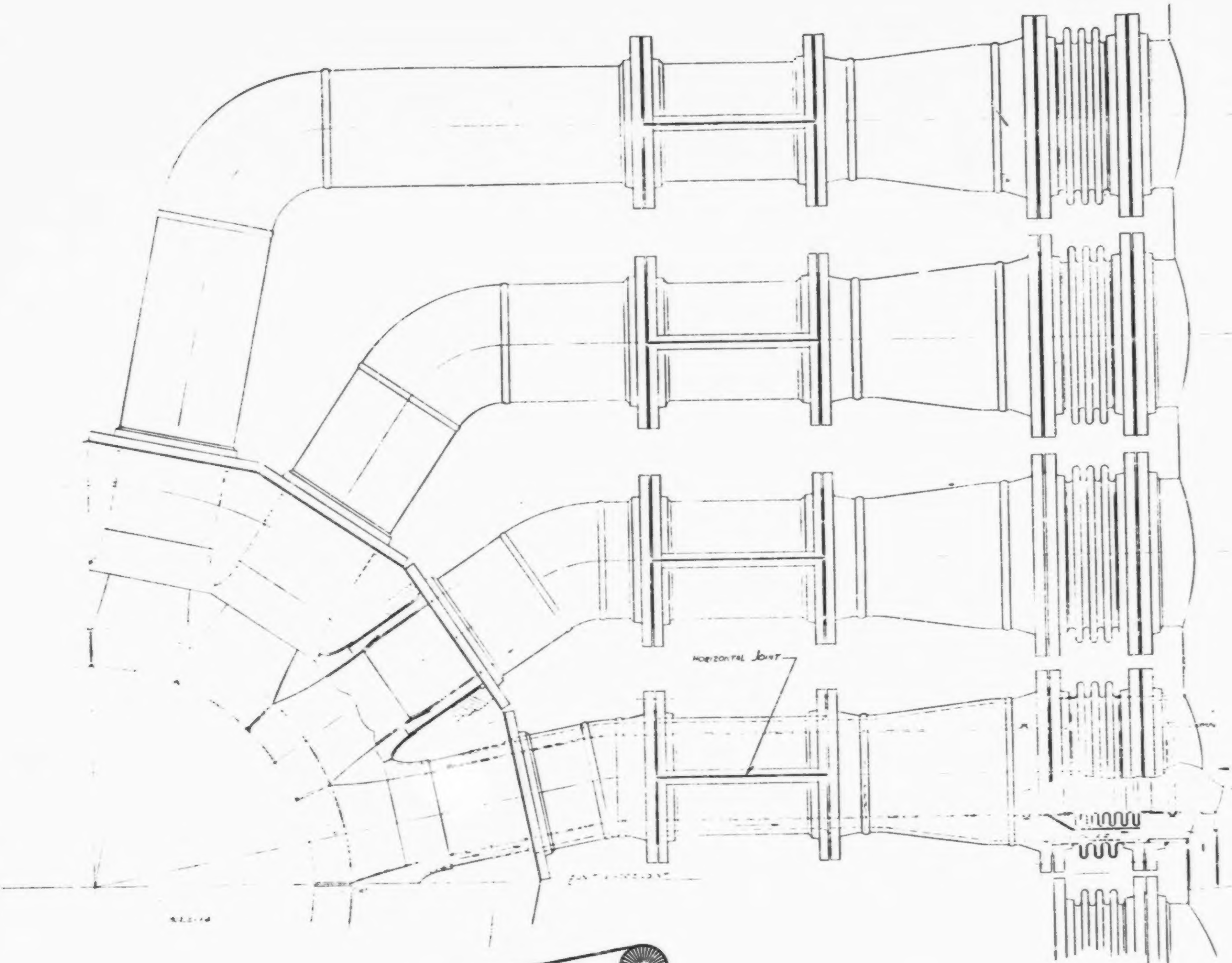

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Fig. 1-120—Detail of
modified W5

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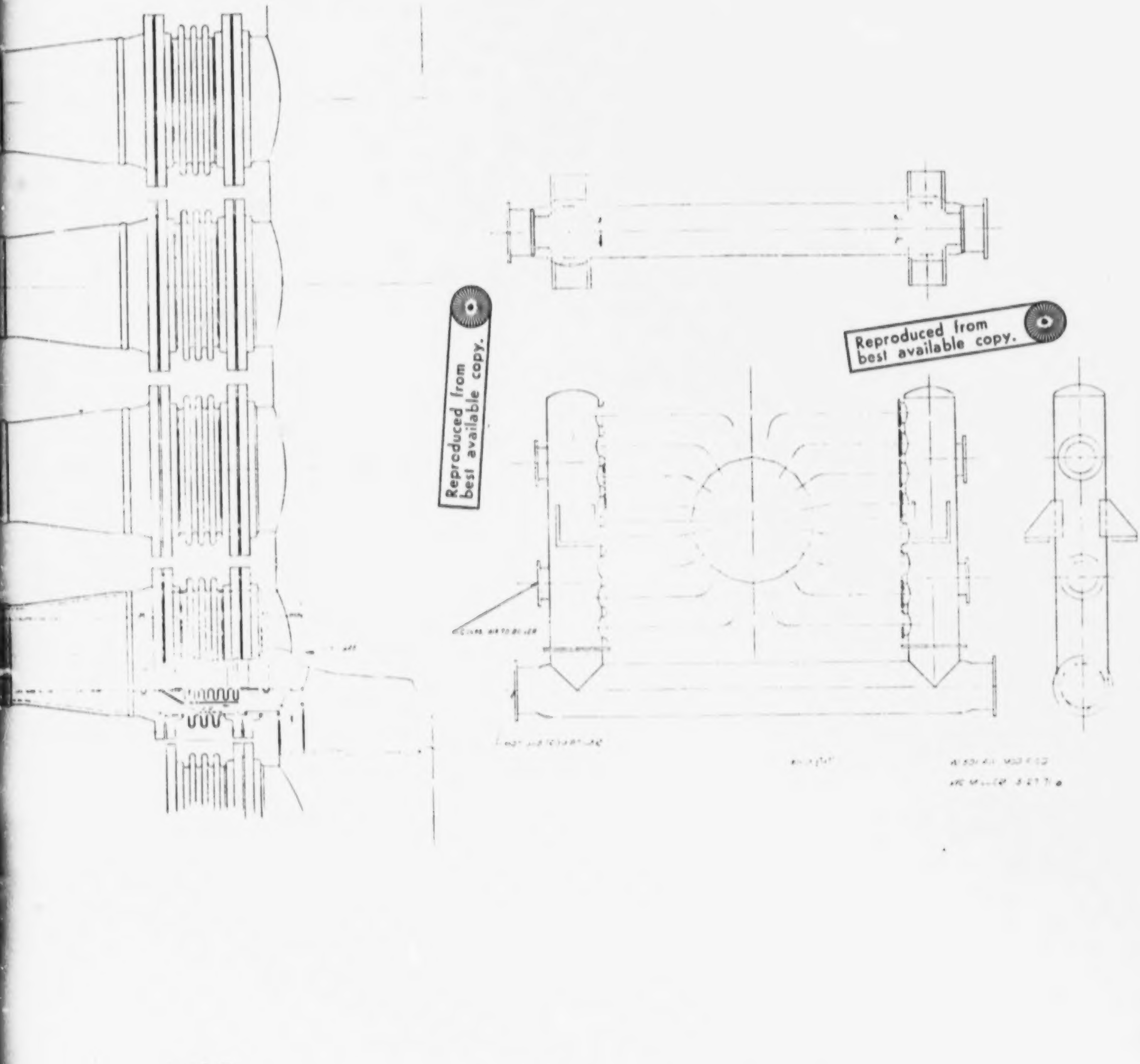

Fig. 1-120—Detail of piping arrangement for modified W501 gas turbine

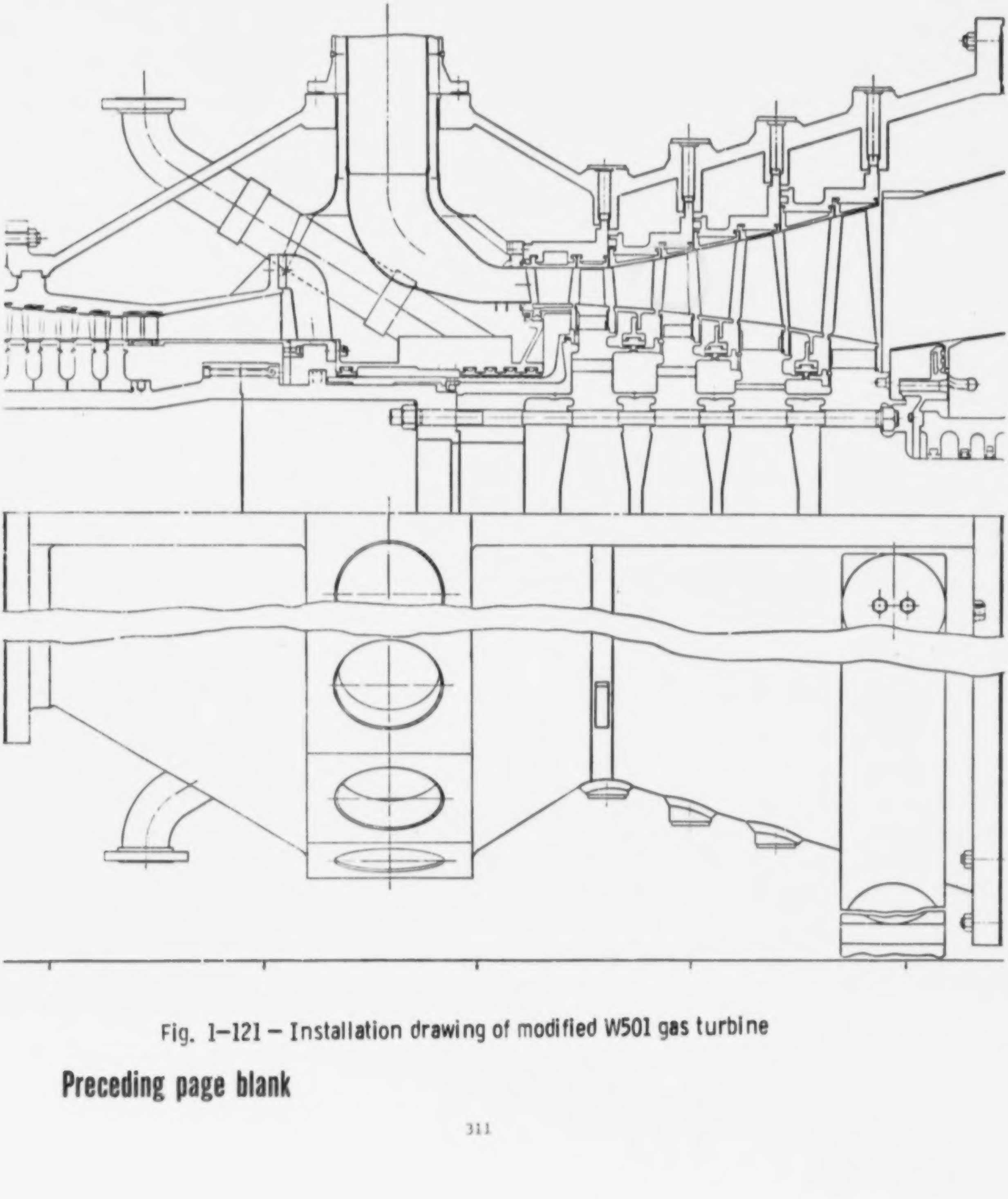

Fig. 1–121 – Installation drawing of modified W501 gas turbine

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Insulated crossover pipe for hot gases		43,600
	Total	$123,700
	Round off to	$125,000

Steam Turbines. The steam conditions assumed for the 318 MW plant are 2400 psig/1000°F/1000°F and 3 in. Hg back pressure. The steam turbine for this plant would be a 3600 RPM tandem compound unit with high-pressure, intermediate-pressure, and low-pressure components. The low-pressure end would be four-flow with 23-inch blades. All four boiler modules would serve one steam turbine with a rating of 265 MW - 300 MVA.

Stack Gas Coolers

A conventional heat recovery system is used to recover the sensible heat from the gas leaving the gas turbines. These stack gas coolers are used to preheat the feedwater.

Exhaust gas leaves the turbine at approximately 830°F and is cooled to 525°F in an upper stack gas cooler while heating 480°F feedwater from the last feedwater heater stages to 578°F. The gas then passes through a lower stack gas cooler where it is cooled to 275°F. The lower stack gas cooler heats feedwater from the feedwater heaters associated with the low-pressure steam turbine section from 225°F to 480°F.

Modular designs and costs were prepared by vendors. The cost, including transition ducting with the gas turbines, is $4.1/kw for both a 318 MW and 635 MW plant. Design specification and questions are presented in Appendix N.

Operation and Controls

The steam flow circuits of the high-pressure boiler are very similar to those of a typical once-through unit. The modular boiler arrangement with independently fired fluidized beds makes the operation of the system more complex, but the basic characteristics of once-through operation are not changed.

In a conventional coal-fired boiler the practice is to operate the furnace at the minimum amount of excess air consistent with good combustion efficiency to minimize stack losses. In a fluid bed boiler, constraints are imposed on the combustion airflow by the hydrodynamics of the bed. In the combined cycle plant, further constraints are placed on combustion airflow by the gas turbine compressor if the power turbine which drives the generator is not separate from the compressor turbine. With a constant speed compressor the airflow can be varied by only about one-third of the boiler capacity.

In some aspects of the steam generator operation, the fluidized bed boiler has greater flexibility than the conventional pulverized coal-fired boiler. The separation of the steam generator functions into separately fired beds is highly advantageous since it permits specific control of heat input to each part of the water/steam circuitry.

Start Up. In starting up the 318 MW high-pressure fluidized bed plant one gas turbine module is started up at a time, and one boiler module is started up at a time.

As shown in Figure 1-122, each gas turbine module is equipped with a by-pass line around the two boiler modules. During start-up about one-half of the compressor airflow would be diverted through this by-pass line where an oil-fired start-up combustor is located. The combustor is ignited and the gas turbine is accelerated to the self-sustaining speed of about 1630 RPM. Here the starting motor is cut out and the unit accelerates to the idle conditions of 3600 RPM and about 1000°F turbine inlet temperature. At this point the air temperature from the gas turbine compressor is about 575°F.

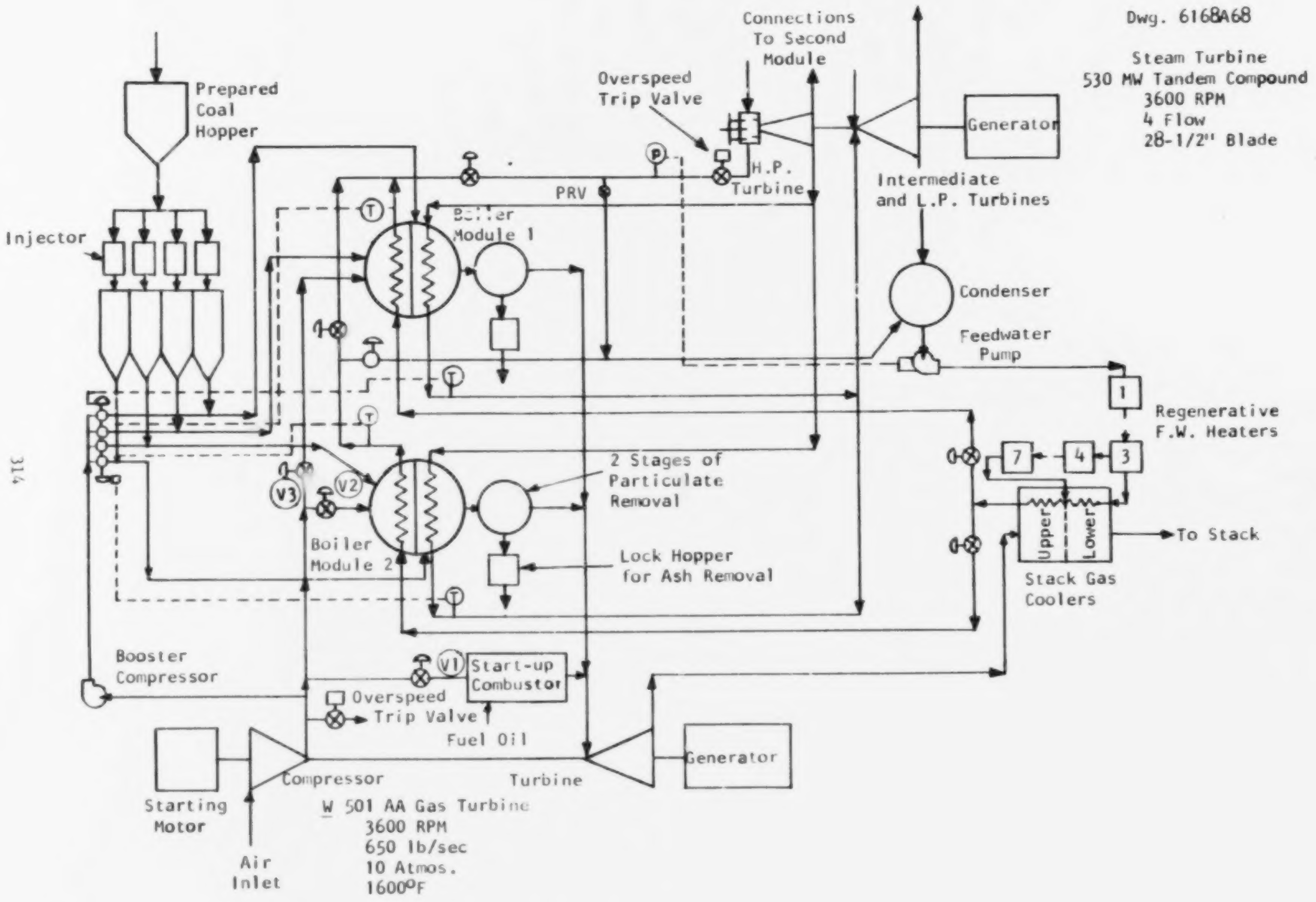

Fig. I-122 — Flow diagram for plant start-up

The start-up of the first boiler module in the plant is different from that of the later modules since there is no steam available for preheating the beds. All beds are equipped with oil-fired igniters to heat the beds to the ignition point of the coal, which is assumed to be 750°F. Water is circulated through the pre-evaporator bed, the wall tubes, and the first superheater. From there it goes to the flash tank and then to the condenser. The pre-evaporator and the first superheater bed depths are reduced to three feet, preheated to the ignition temperature, and coal feeding is started. Once ignition is established, dolomite is fed into the beds to bring the bed depths up to full capacity in a period of about one hour. The pressure and temperature are built up in this circulating loop with the steam from the flash tank going to the condenser and to the second superheater tubes. When the gas temperature from the boiler module is sufficiently high, the start-up combustor is turned off. Next, the second superheater bed is ignited; and, when the steam conditions are adequate, the turbine is warmed and rolled. Then the reheat and carbon burn-up cell beds are ignited. The steam pressure is increased until it reaches design value with an output of about 10% full plant load or 40% of module rating. The turbine is then put on governor control.

The other boiler modules are started up as needed by preheating the beds with steam from the first module. Since feedwater is now at a temperature of about 500°F, the start-up times for the second, third, and fourth boiler modules are much shorter than for the first. Detailed descriptions of the boiler start-up procedure are given in Appendix H by Foster Wheeler and in Appendix J by United Engineers and Constructors.

Load Control. The specified turn-down requirement for this plant is 4:1. It has been shown that a modular boiler system with each module having a capability of 50% turn-down is required to meet this specification with no discontinuities. With the arrangement of two gas turbine modules and four boiler modules which was selected for this preliminary design study, a 4:1 turn-down is attainable. However, a minimum primary bed temperature as high as 1400°F may be required for adequate sulfur removal, whereas a bed temperature of about 1325°F is required to attain 25% plant capacity with one gas turbine module and two boiler modules. Therefore, there may be a discontinuity in steady state operation between 25% and 31% of plant capacity.

Tables 1-42 and 1-43 list the operating parameters of the individual fluidized beds for design and 70% plant load. This shows that the bed temperatures vary significantly at part load because of a shift in the distribution of heat input among the boiler functions as load changes, and that the superficial velocities do not vary substantially with load change.

Table 1-44 lists the steam generator parameters for design and 70% plant load.

The specification for load change rate for this plant is 5% per minute. Preliminary analysis has indicated that this specification can probably be met. However, the beds do have a substantial thermal inertia, and the residence time of the fuel in the bed is relatively high. Therefore, a comprehensive analysis of the load change rate capability is needed.

Shut-down. The normal shut-down procedure for the high-pressure boiler is basically a reversal of the start-up procedure. Fluidizing air to the beds will be maintained after fuel feed has stopped in order to purge and cool the beds. If it is necessary to keep the bed on hot standby, the flow of feedwater to the beds would be maintained.

Two situations would lead to emergency shut-down. One is loss of load on the gas turbine, and the other is loss of load on the steam turbine. An emergency relief valve is provided in the air by-pass line to handle loss of load on the gas turbine (see Figure 1-122). If such a loss were to occur, the fuel flow to the beds would automatically be stopped, and the relief valve would open to divert flow from the gas turbine.

Emergency shut-down due to loss of load on the steam turbine would be accomplished in the same manner as in a conventional steam

TABLE 1-42

FLUIDIZED BED PARAMETERS
DESIGN LOAD
(1 MODULE)
GAS TEMPERATURE LEAVING BOILER(a) - 1650°F

	FUEL FLOW (LB/HR)	AIR FLOW (LB/HR)	FLUE GAS (LB/HR)	SUPERFICIAL VELOCITY	BED TEMPERATURE
Pre-evaporator	17,680	192,100	206,200	9.0	1,750
1st Superheater	13,100	142,700	153,700	6.8	1,750
2nd Superheater	10,740	116,900	125,900	5.6	1,750
Reheater	12,390	134,800	145,200	6.4	1,750
CBC	6,494	42,600	44,660	6.1	2,000

(a) Includes CBC Gas

(b) 31.7% carbon
63.3% ash
5.0% dolomite

TABLE 1-43

FLUIDIZED BED PARAMETERS
70% PLANT LOAD
(1 MODULE)
GAS TEMPERATURE LEAVING BOILER(a) - 1400°F

	FUEL FLOW (LB/HR)	AIR FLOW (LB/HR)	FLUE GAS (LB/HR)	SUPERFICIAL VELOCITY	BED TEMPERATURE
Pre-evaporator	12,600	199,200	209,200	8.7	1,470
1st Superheater	8,950	141,200	148,600	6.3	1,470
2nd Superheater	7,370	116,300	122,400	5.2	1,470
Reheater	9,680	152,200	160,800	7.0	1,530
CBC	4,650	20,200	21,690	3.2	2,000

(a) Includes CBC gas

(b) 31.2% carbon
62.1% ash
6.7% dolomite

TABLE 1-44

STEAM GENERATOR PERFORMANCE SUMMARY

(FOR TOTAL PLANT)

	DESIGN LOAD	70% PLANT LOAD
Fuel flow -- M lb/hr	215	154
Air flow -- M lb/hr	2,506	2,506
Flue gas flow -- M lb/hr	2.705	2,650
Feedwater inlet temperature -- °F	578	547
Feedwater inlet pressure -- psig	2,824	2,650
Main steam outlet temperature -- °F	1,000	1,000
Main steam outlet pressure -- psig	2,500	2,444
Main steam flow -- M lb/hr	1,727	1,140
Reheat inlet temperature -- °F	650	584
Reheat inlet pressure -- psig	600	405
Reheat outlet temperature -- °F	1,000	1,000
Reheat outlet pressure -- psig	580	391
Reheat flow -- M lb/hr	1,644	1,090
Flue gas exit temperature -- °F	1.650	1,400
Flue gas pressure drop -- psig	4.8	4.8
Fuel/air ratio	0.0856	0.0615

plant. Flow would be shut off to the turbine by an emergency trip valve, and the steam would be diverted to the condenser. Fuel flow to the beds would be shut off automatically, and airflow would be maintained to purge and cool the beds. If possible, feedwater flow would be maintained until the beds had cooled sufficiently so that tubes would not be overheated.

Control and Instrumentation

A composite flow diagram for the high-pressure fluidized bed boiler system is shown in Figure 1-123. The control system for this plant is designed to perform the following functions:(1)

- Maintain steam header pressure to the turbine at the desired value by regulation of the turbine governor valve
- Maintain the steam header temperature at the turbine at the desired value by regulation of the fuel input to the beds
- The boiler feedwater flow is set by the load demand from the dispatcher.

Many of the plant subsystems are automated, e.g., that which maintains solid levels in feeder bins. Other functions are manually controlled, e.g., the sorbent recycling rate to the beds where a monitor sounds an alarm when the recycle rate is too low or too high. The Foster Wheeler report (Appendix H) gives complete details on the controls for the boiler and the boiler auxiliaries and lists the instrumentation required for control and monitoring.

Plant Layout

United Engineers and Constructors prepared a plant layout for a 635 MW high-pressure fluidized bed boiler system (see Figure 1-124). A hypothetical site was assumed, located in the eastern United States on a large river - the source of once-through cooling water. Access is provided to a state highway by an existing secondary road. Rail access

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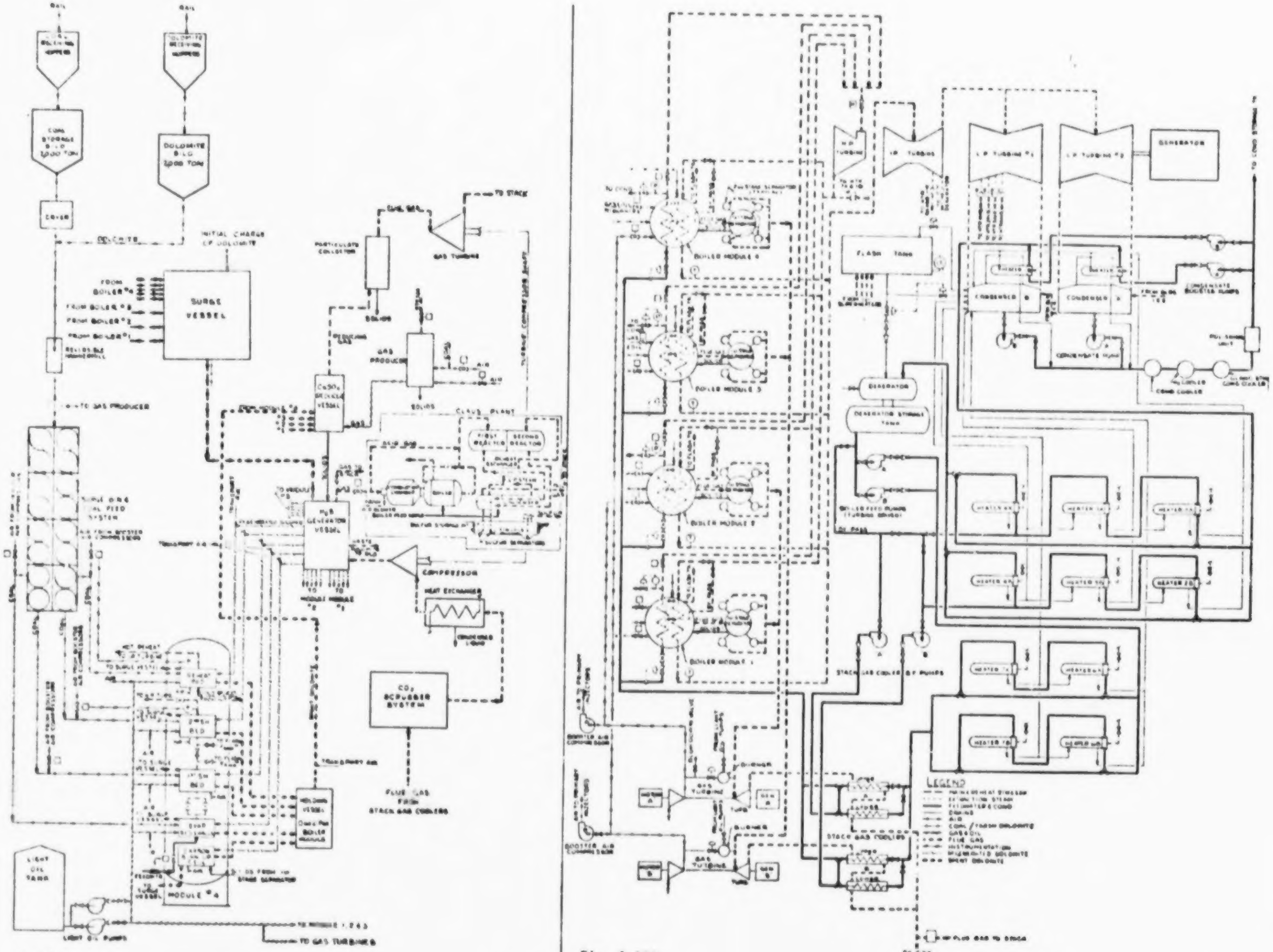
(1) More details are given in the United Engineers and Constructors report (see Appendix J).



635 MW FLUIDIZED BED
BOILER COMBINED CYCLE PLANT
COMPOSITE FLOW DIAGRAM
united engineers

Fig. 1-123

321-1

321-2

321-3

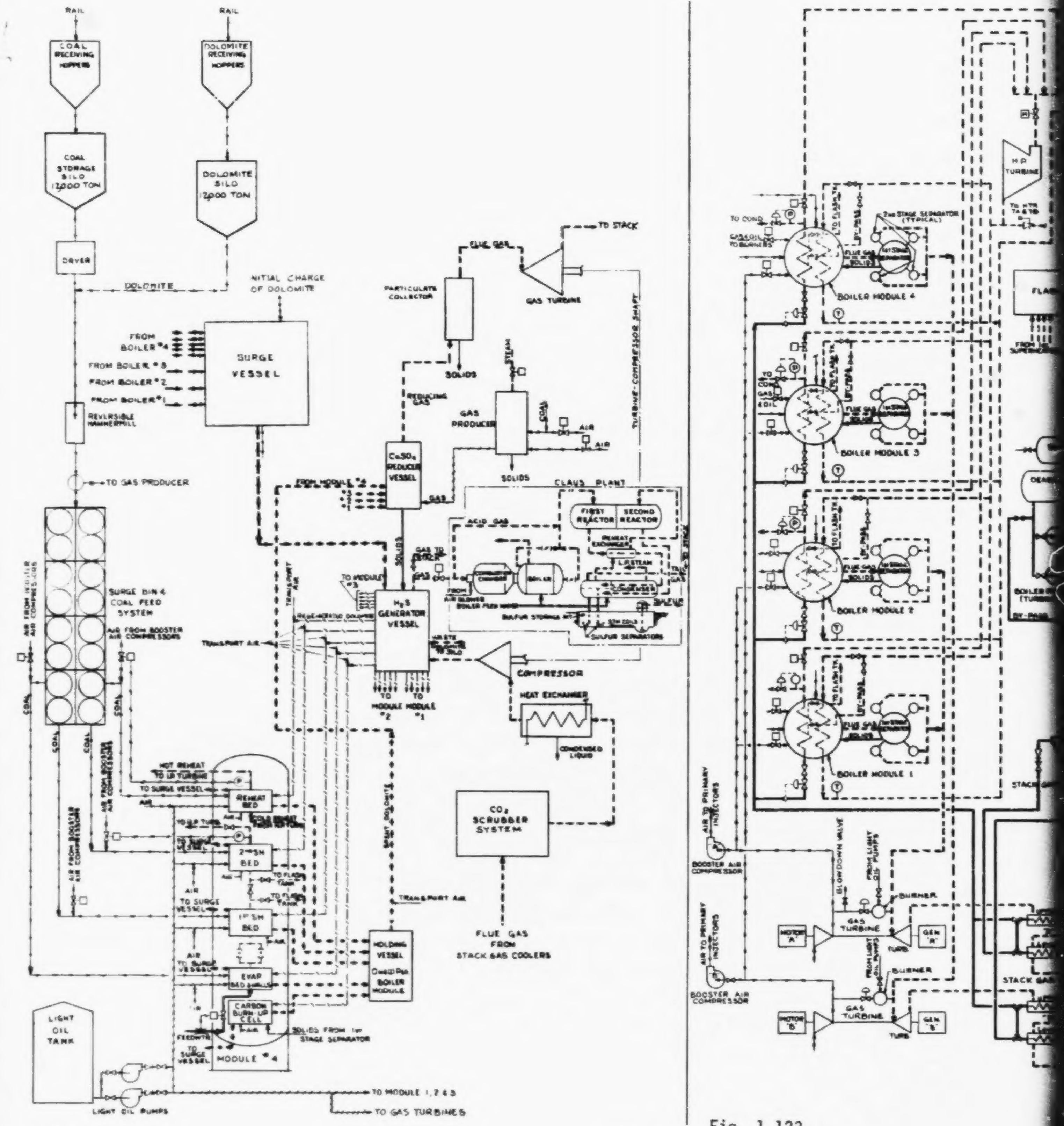
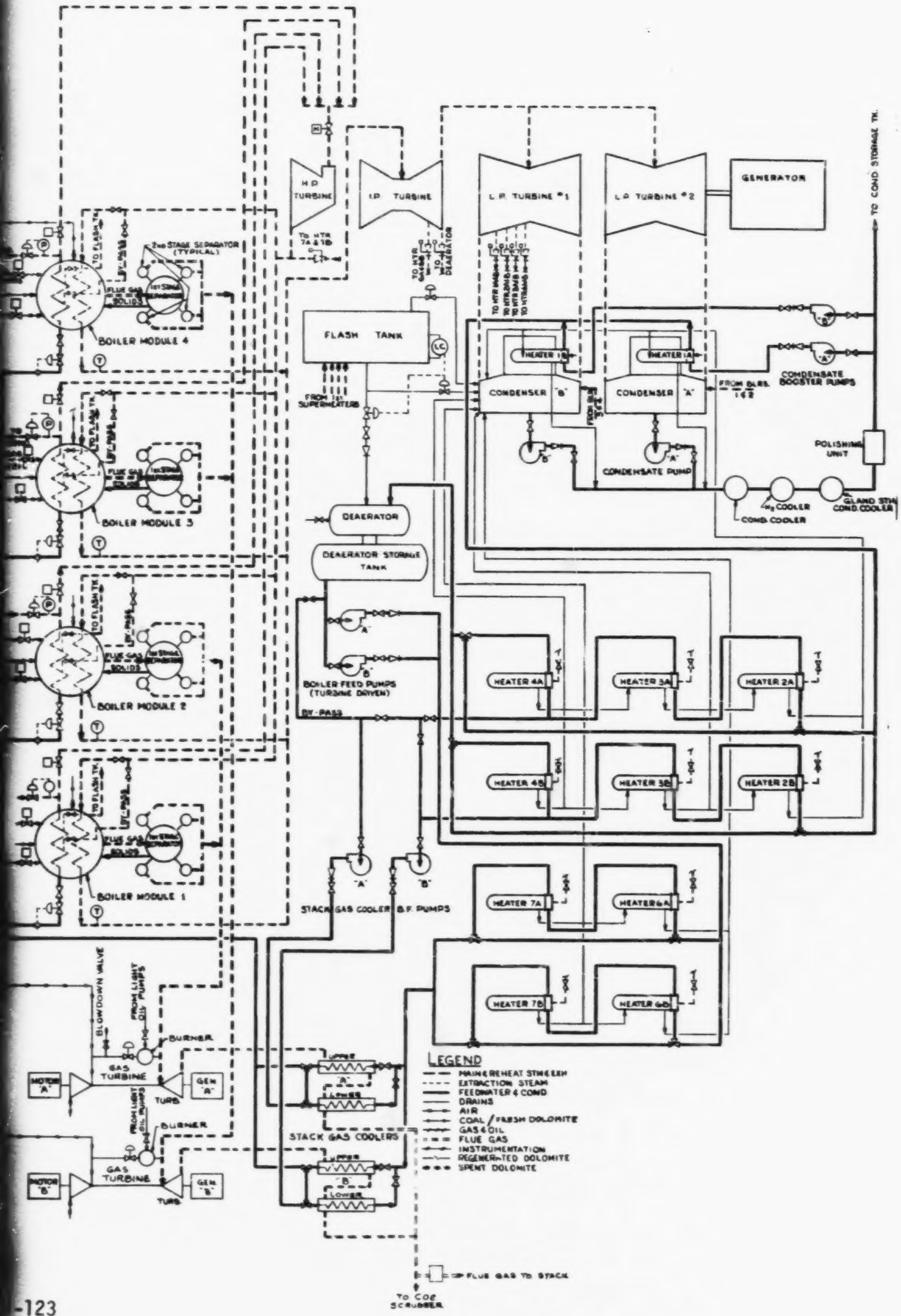


Fig. 1-123

321.1

321.2



635 MW FLUIDIZED BED
BOILER COMBINED CYCLE PLANT
COMPOSITE FLOW DIAGRAM

united engineers & constructors inc.

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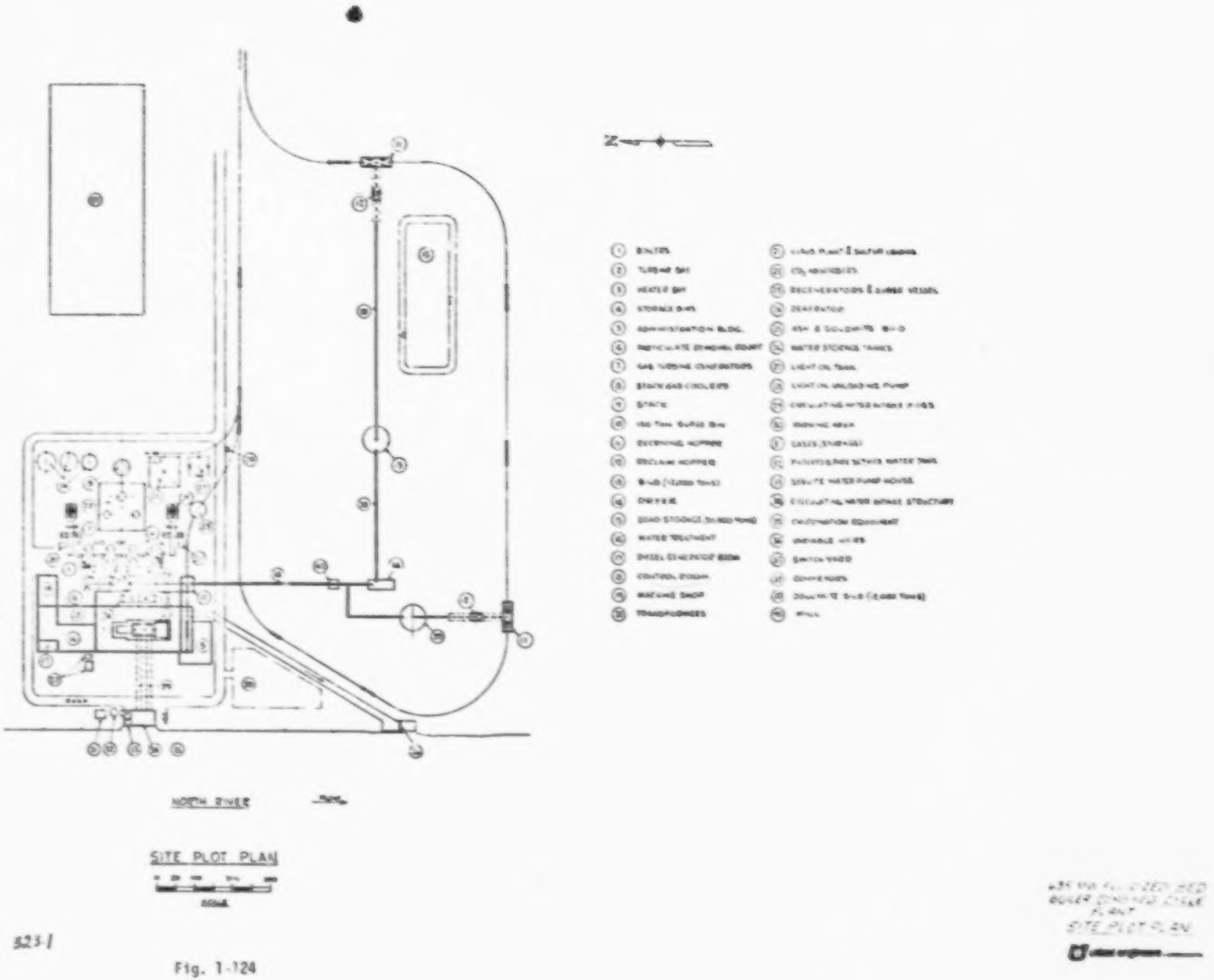

Fig. 1-124

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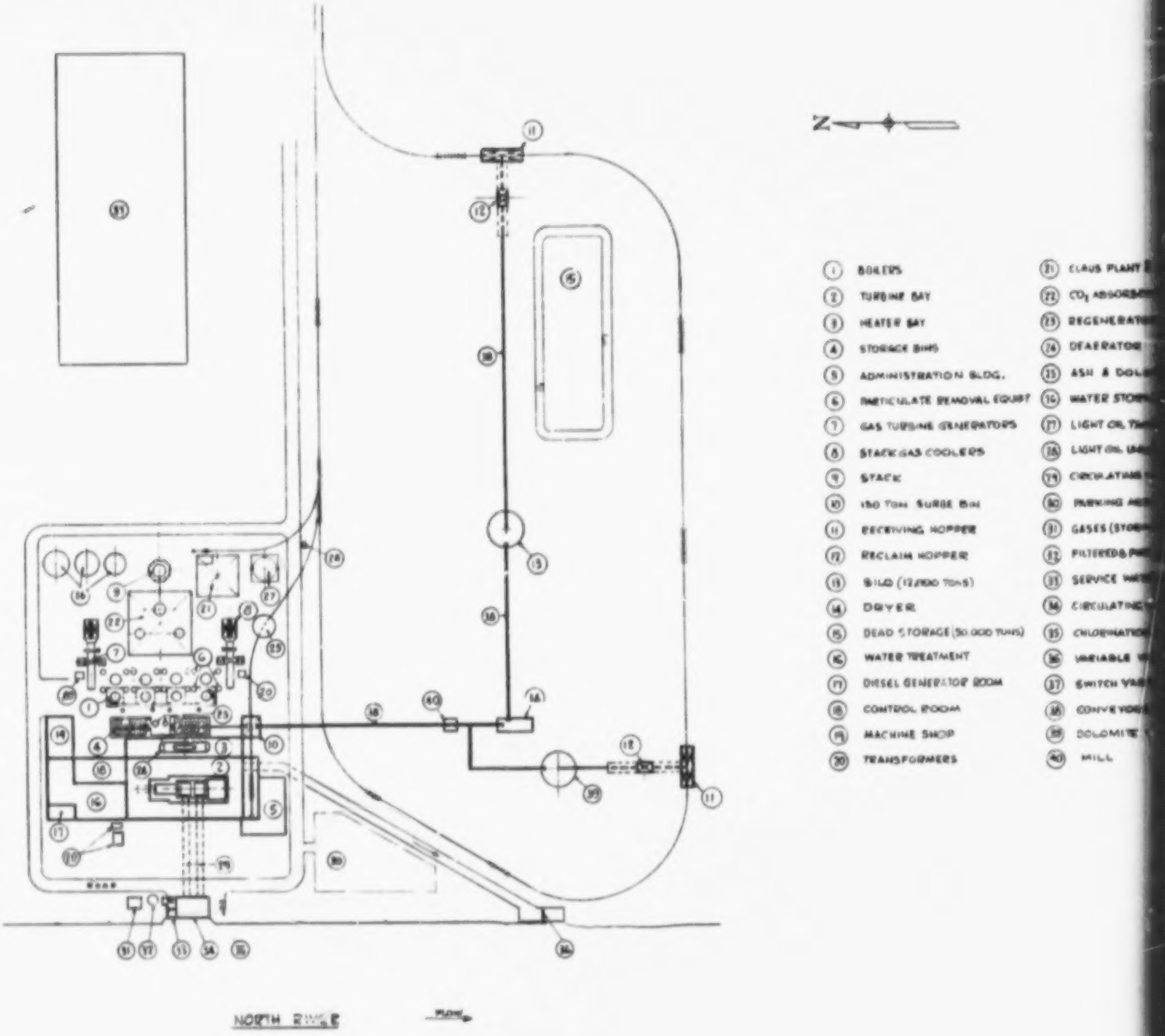
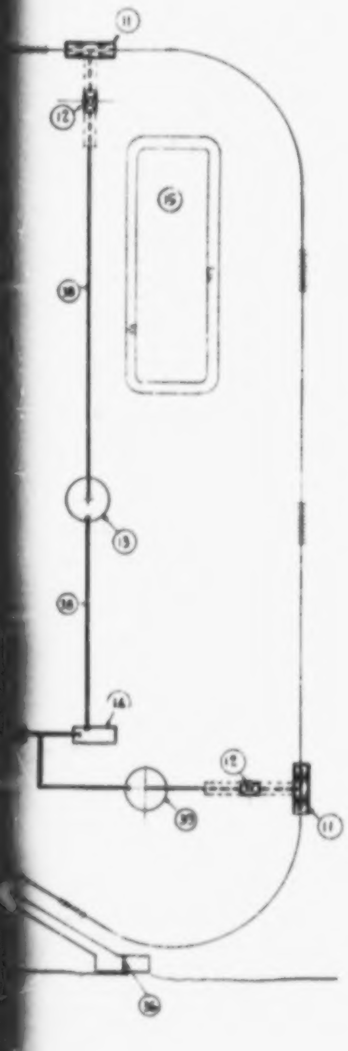


SITE PLOT PLAN

Fig. 1-124



① BOILERS
② TURBINE BAY
③ HEATER BAY
④ STORAGE BINS
⑤ ADMINISTRATION BLDG.
⑥ PARTICULATE REMOVAL EQUIPT
⑦ GAS TURBINE GENERATORS
⑧ STACK GAS COOLERS
⑨ STACK
⑩ 150 TON SURGE BIN
⑪ RECEIVING HOPPER
⑫ RECLAIM HOPPER
⑬ SILO (12,000 TONS)
⑭ DRYER
⑮ DEAD STORAGE (50,000 TONS)
⑯ WATER TREATMENT
⑰ DIESEL GENERATOR ROOM
⑱ CONTROL ROOM
⑲ MACHINE SHOP
⑳ TRANSFORMERS
㉑ CLAUS PLANT & SULFUR LOADING
㉒ CO_2 ABSORBERS
㉓ REGENERATORS & SURGE VESSEL
㉔ DEAERATOR
㉕ ASH & DOLOMITE SILO
㉖ WATER STORAGE TANKS
㉗ LIGHT OIL TANK
㉘ LIGHT OIL UNLOADING PUMP
㉙ CIRCULATING WATER INTAKE PIPES
㉚ PARKING AREA
㉛ GASES (STORAGE)
㉜ FILTERED & FIRE SERVICE WATER TANK
㉝ SERVICE WATER PUMP HOUSE
㉞ CIRCULATING WATER INTAKE STRUCTURE
㉟ CHLORINATION EQUIPMENT
㊱ VARIABLE WEIRS
㊲ SWITCH YARD
㊳ CONVEYORS
㊴ DOLOMITE SILO (12,000 TONS)
㊵ MILL

635 MW FLUIDIZED BED
BOILER COMBINED CYCLE
PLANT
SITE PLOT PLAN

united engineers

can be obtained by the construction of a five-mile spur. Other drawings prepared by United Engineers and Constructors and included in this report are:

Plan of Main Equipment at Grade Floor, Figure 1-125
Plan of Main Equipment at Operating Floor, Figure 1-126
Elevation of Main Equipment, Figure 1-127
Plan of Main Equipment at Mezzanine Floor, Figure 1-128

Drawings included only in the United Engineers and Constructors report (Appendix J) are:

Isometric of Boiler Modules and Particulate Separator
Details of Gas Piping from Separators to Gas Turbines
Electrical Single Line Diagram
Steam Turbine and Surge Bin Foundations
Steam Turbine and Surge Bin Structures
Steam Turbine Foundations - Plans and Sections.

In order to have a basis for comparing the physical characteristics of the high-pressure fluidized bed combined cycle plant with conventional fossil-fueled plants, United Engineers and Constructors provided drawings of a 1000 MW coal-fired plant (Figures 1-129 - 1-132) and of a 1000 MW oil-fired plant (Figures 1-133 - 1-136).

Plant Performance

Figure 1-102 shows the heat rate versus the load for the 635 MW high-pressure fluidized bed boiler system for the following conditions:

- Back pressure - 3 in Hg
- Combustion efficiency - 89.6%
- Gas turbine system pressure drop - 3%.

The site chosen by United Engineers and Constructors for comparison of energy costs has a supply of once-through cooling water which will give a back pressure of 1-1/2 in. Hg. The estimated combustion efficiency for the high-pressure combined cycle system is 88.6%. The estimated pressure losses in the gas turbine combustor loop at design load are as follows:

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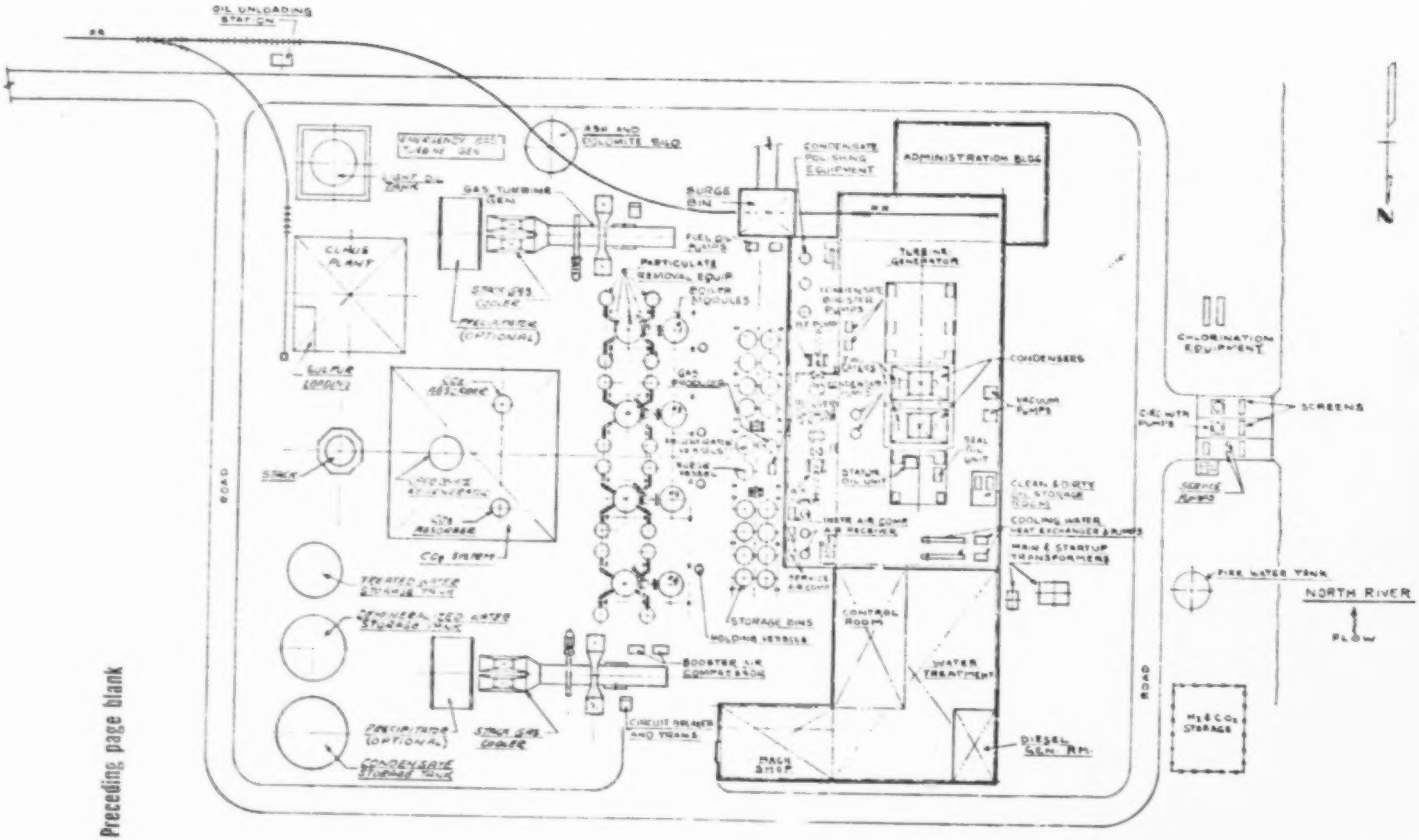


PLAN @ GRADE FL. EL. 18'-0"

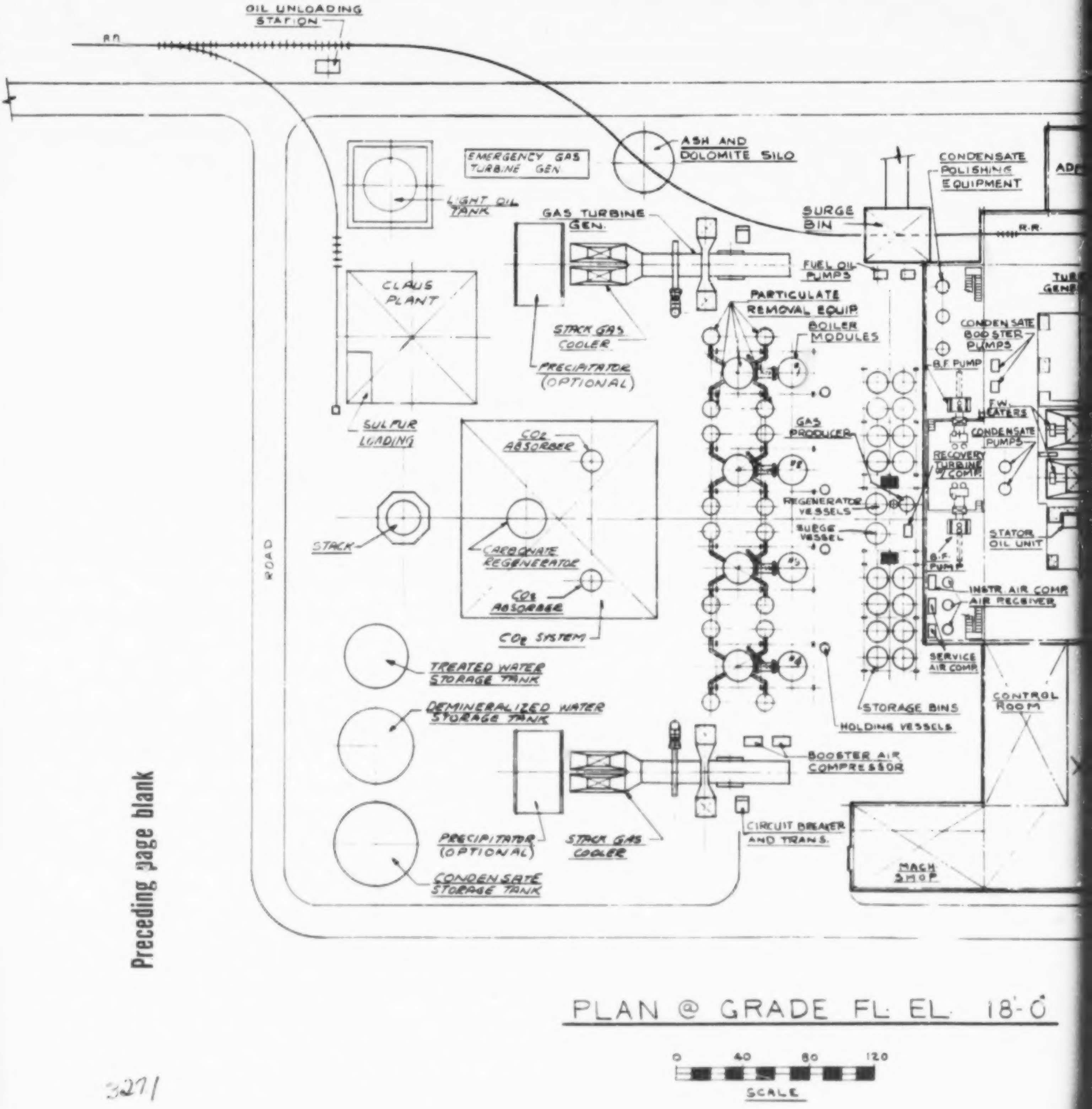
0 40 80 120
SCALE

635 MW FLUIDIZED BED BOILER
COMBINED CYCLE PLANT
PLAN @ GRADE FLOOR - ELEVATION 18'0"
UNITED ENGINEERS & CONSTR INC PHILA, PENNA

327l

Fig. 1-125
327.2

327. 3



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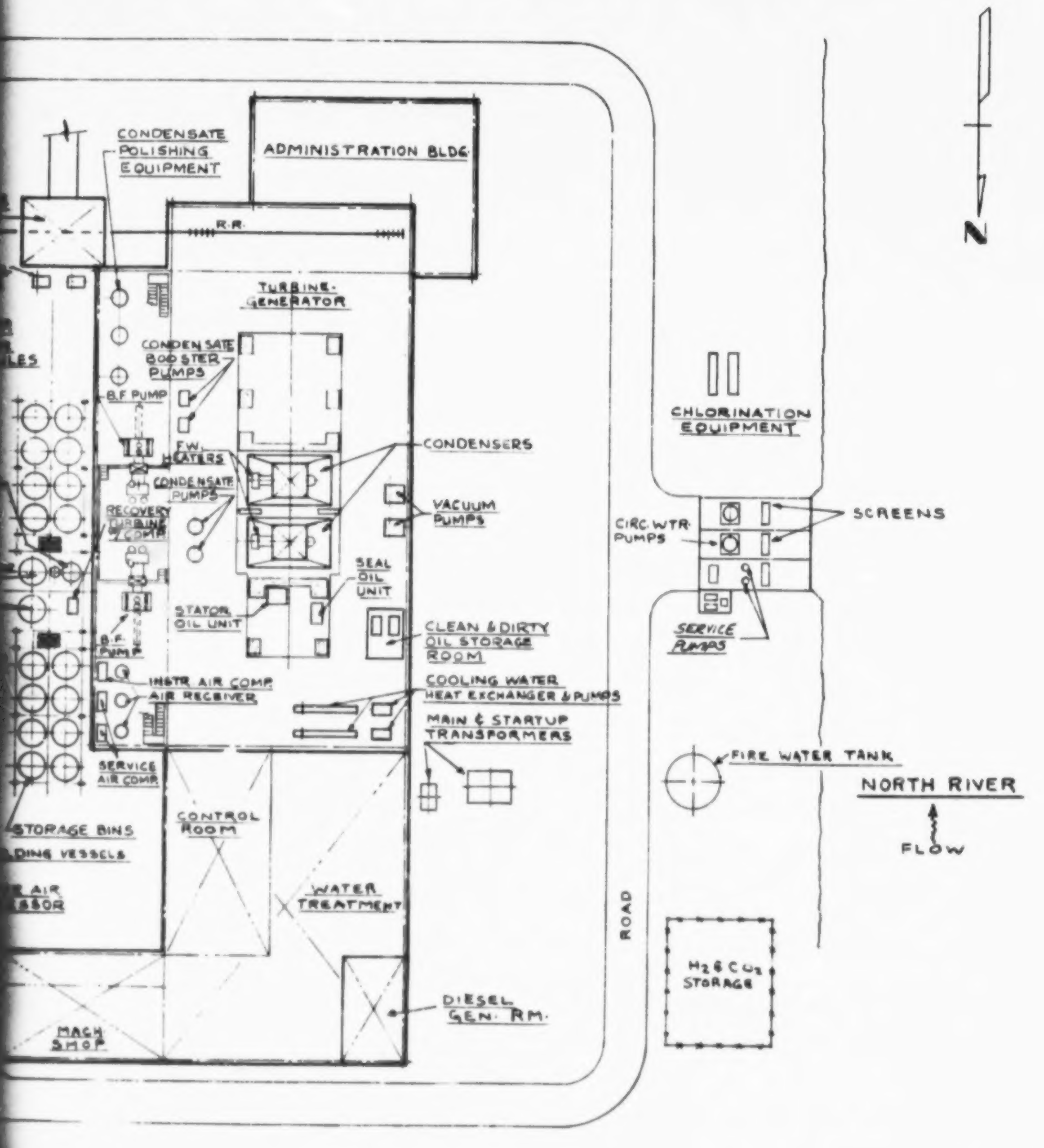
327/

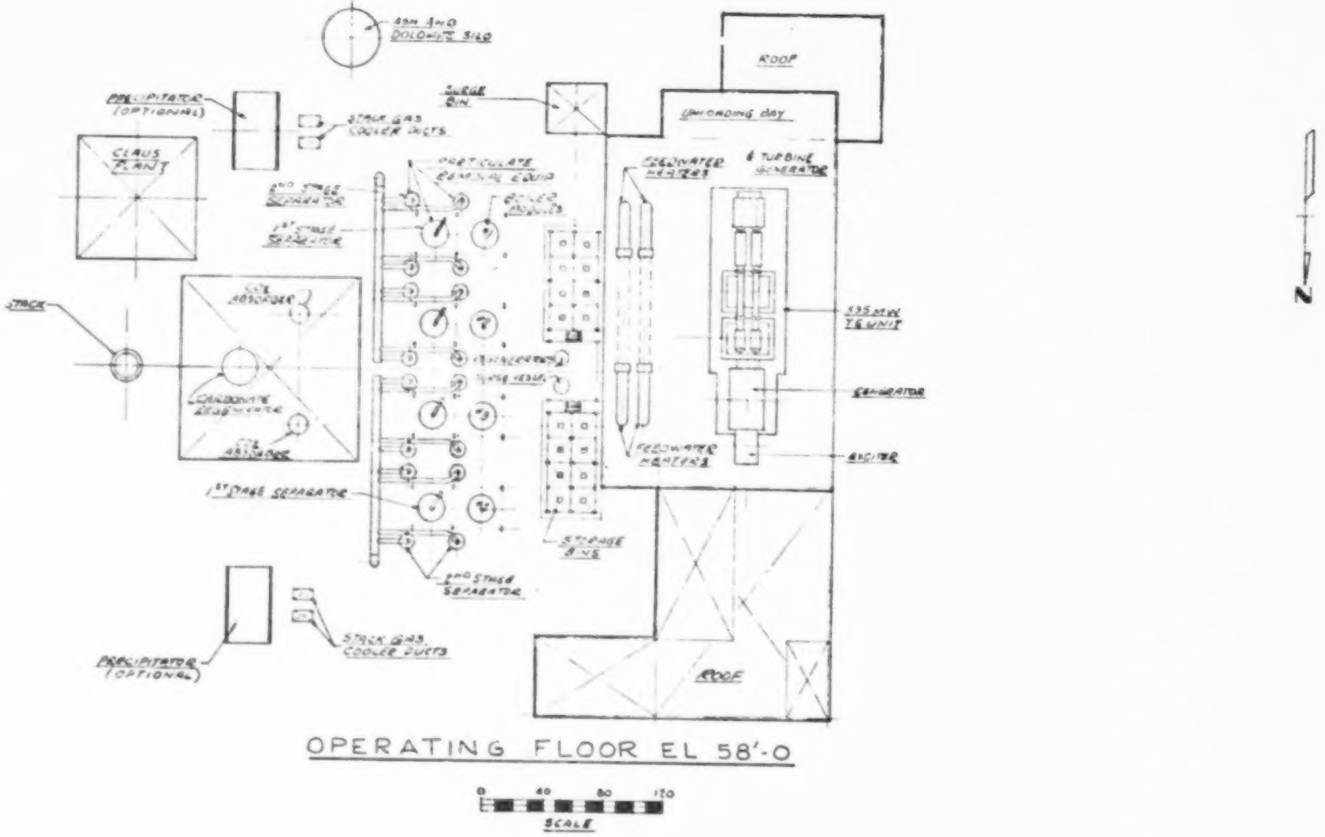
Fig. 1-125

327.2



FL. EL. 18'-0"

635 MW. FLUIDIZED BED BOILER
COMBINED CYCLE PLANT
PLAN @ GRADE FLOOR - ELEVATION 18'0"
UNITED ENGINEERS & CONSTR. INC. PHILA, PENNA.



OPERATING FLOOR EL 58'-0

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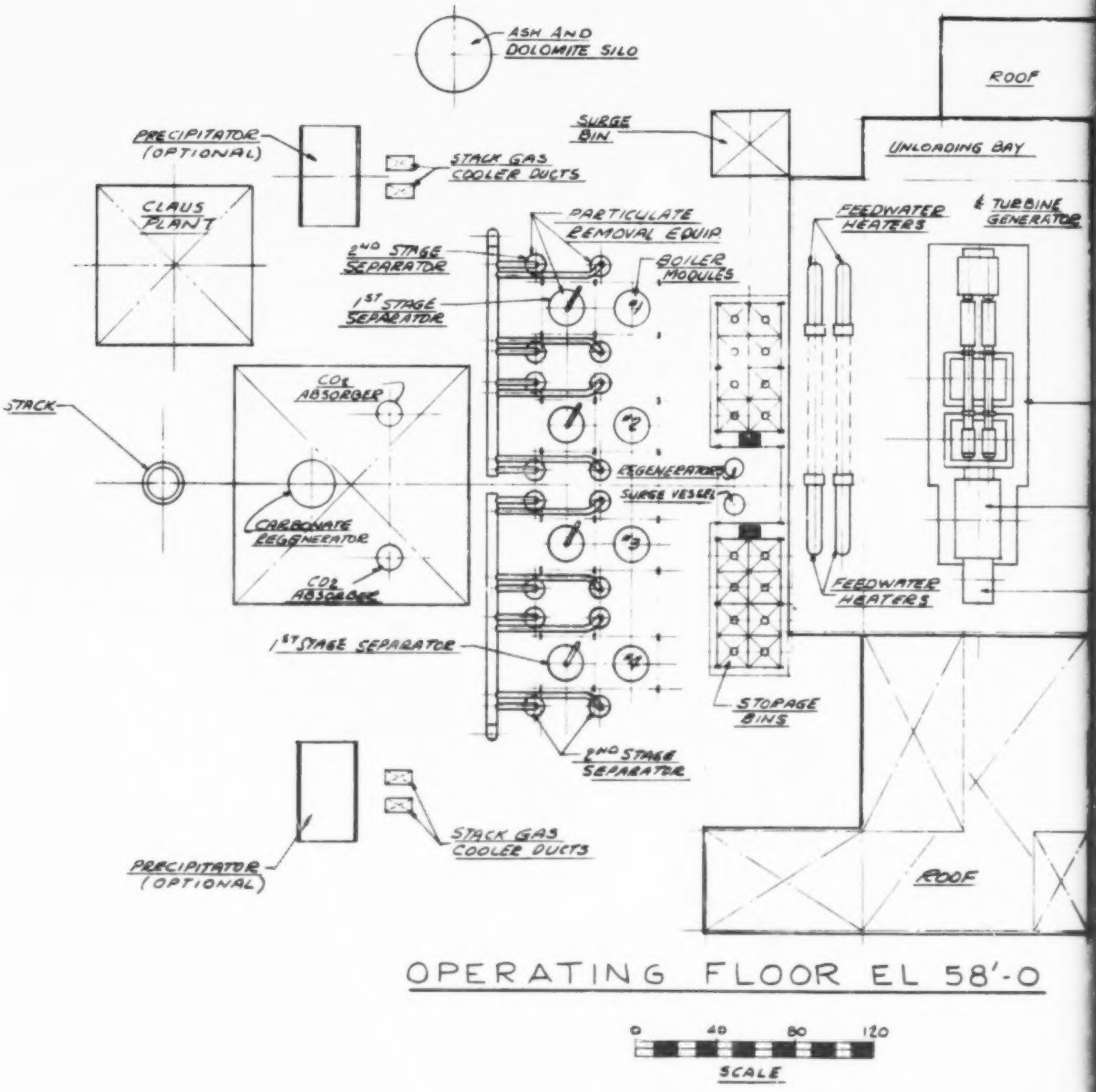
635 MW FLUIDIZED BED BOILER
COMBINED CYCLE PLANT
OPERATING FLOOR PLAN ELEV 58'-0"
UNITED ENGINEERS & CONST'R INC PHILA, PENNA

3291

Fig. 1-126



OPERATING FLOOR EL 58'-0

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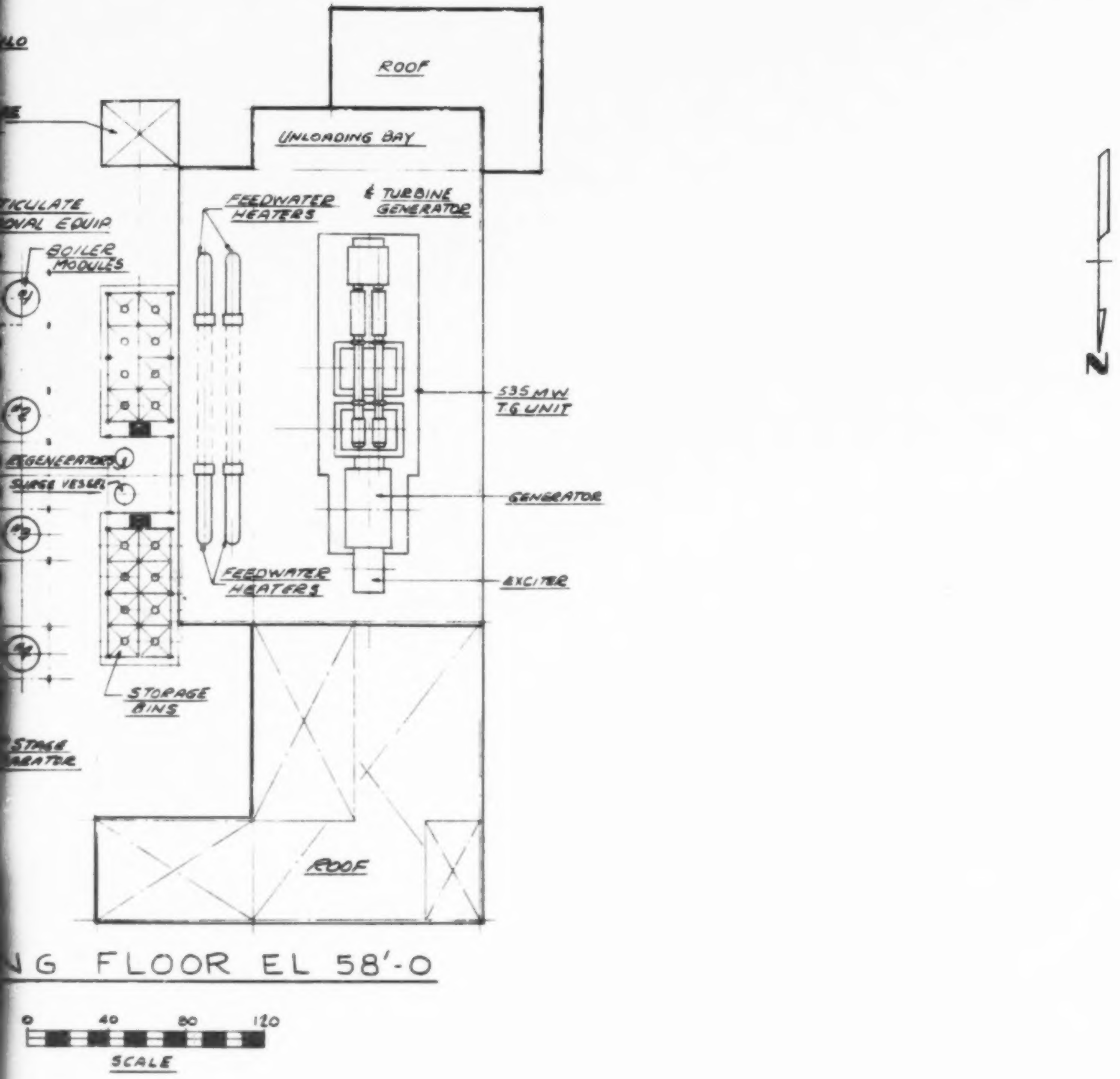
3291

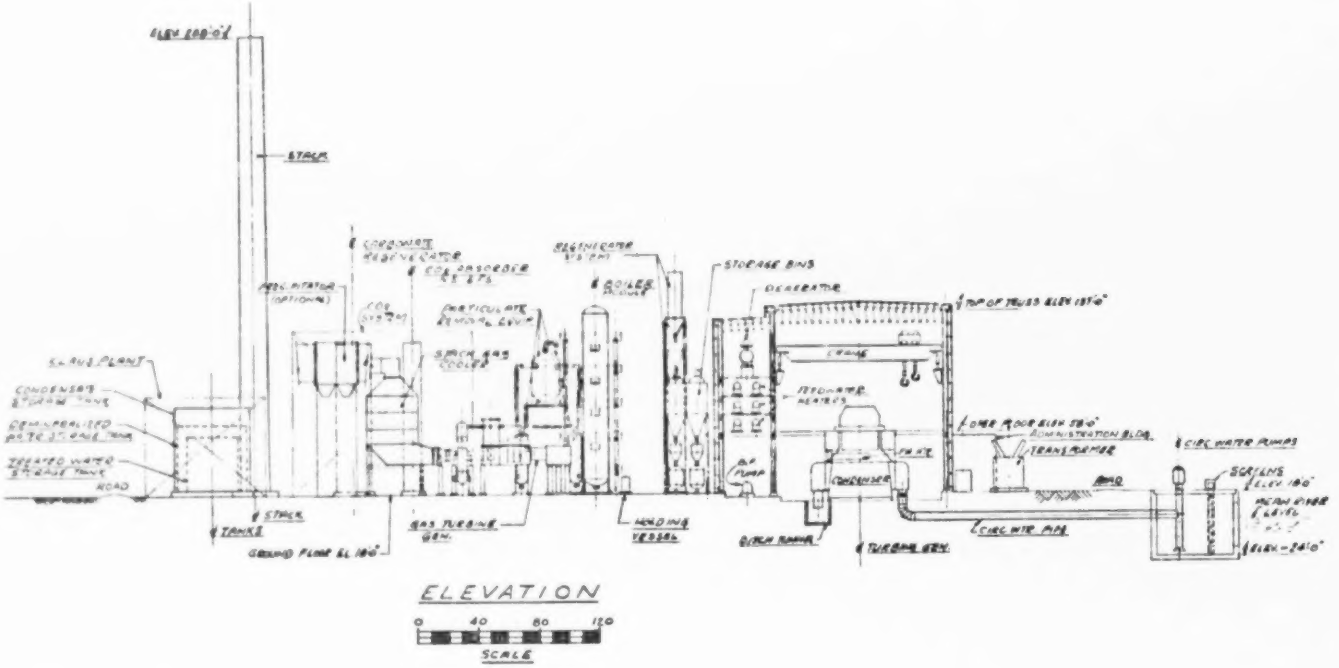
Fig. 1-126



635 M.W. FLUIDIZED BED BOILER
COMBINED CYCLE PLANT
OPERATING FLOOR PLAN ELEV. 58'-0"

UNITED ENGINEERS & CONSTR. INC. PHILA., PENNA.

Fig. 1-126



635 MW FLUIDIZED BED BOILER
COMBINE CYCLE PLANT
ELEVATION
UNITED ENGINEERS & CONSTR INC PHILA, PENNA

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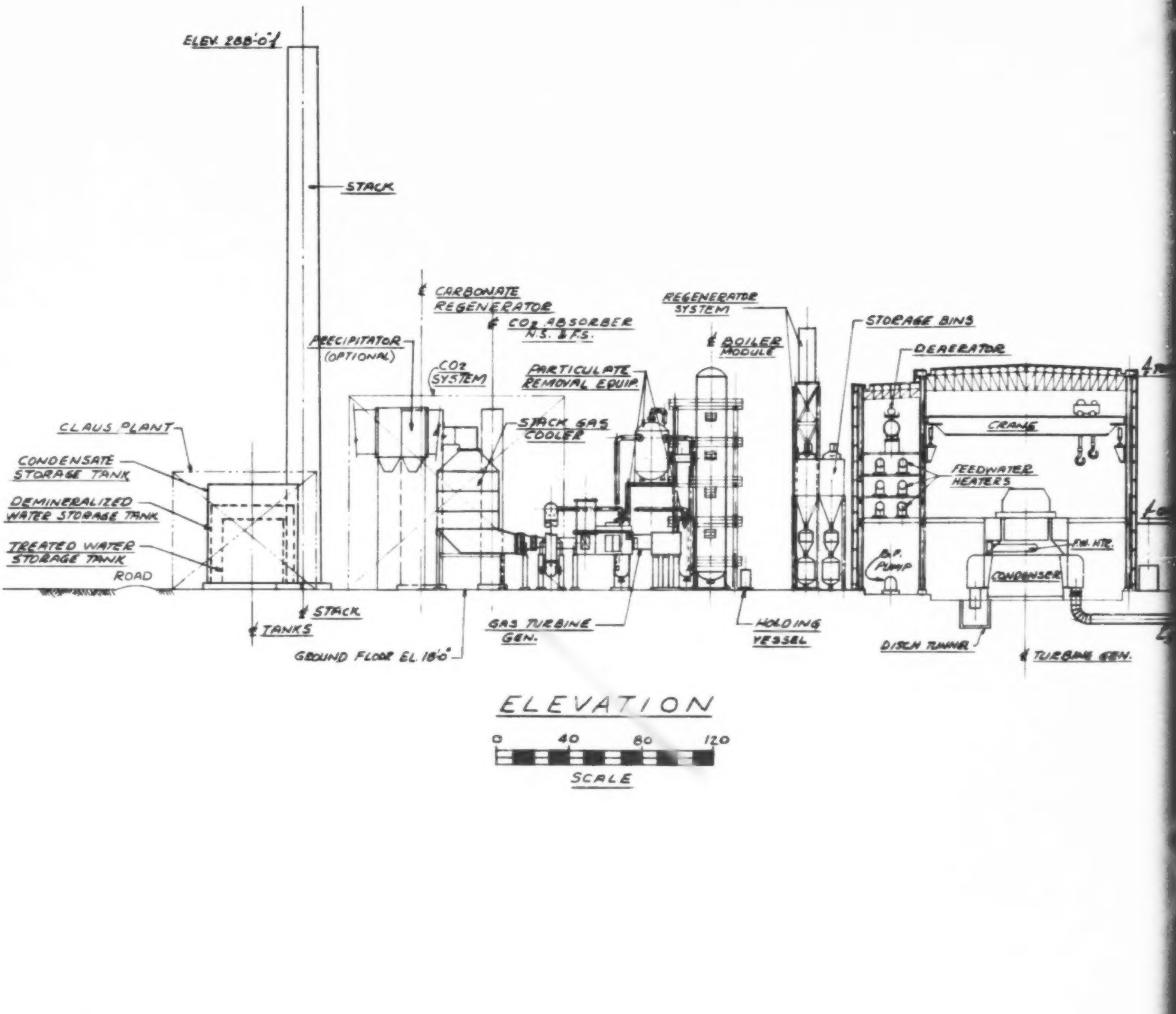
Fig. 1-127

331.1

331.2

331.3



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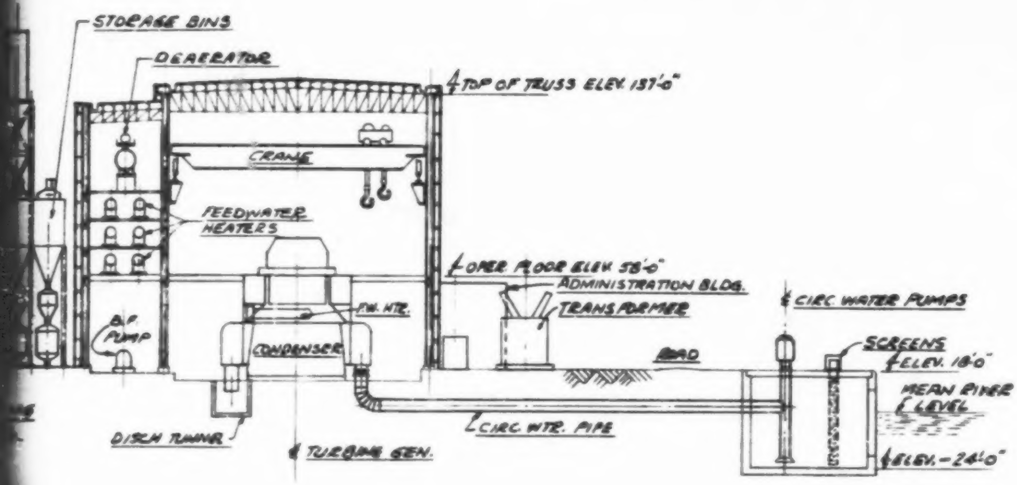
331.1

Fig. 1-127

331.2

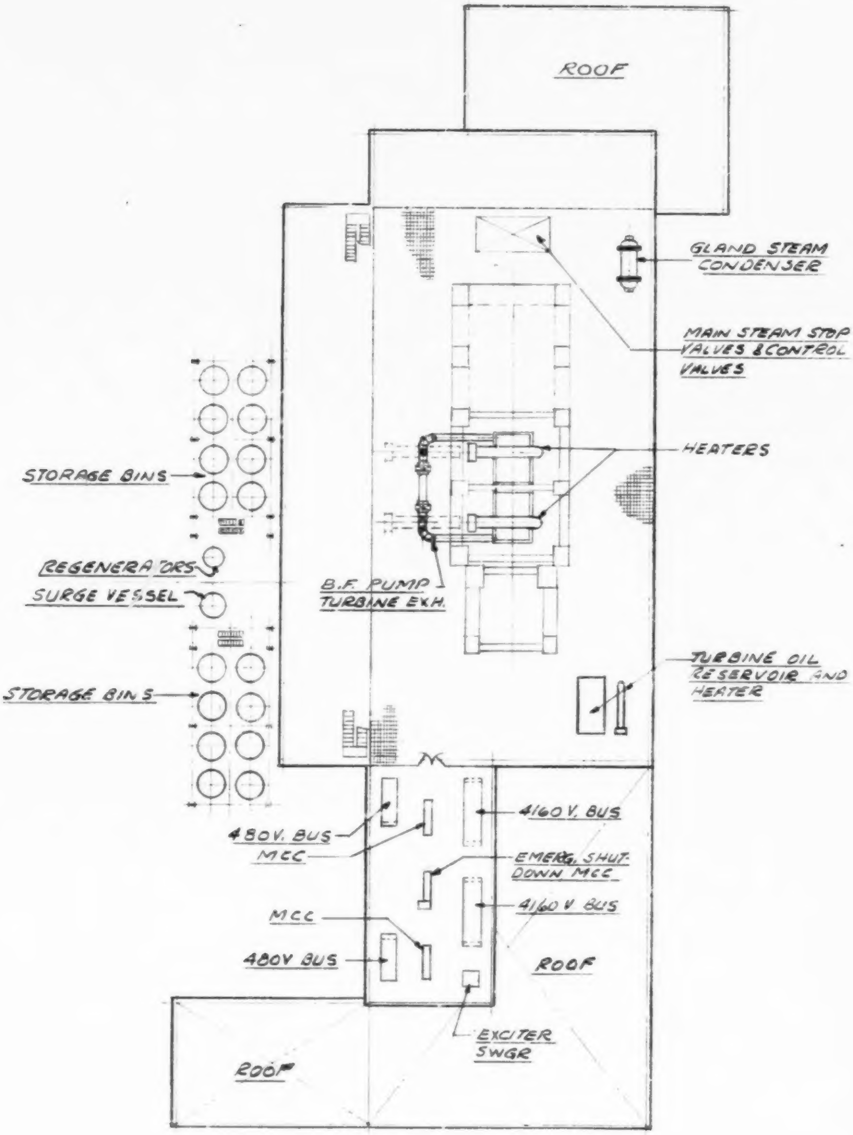


635 M.W. FLUIDIZED BED BOILER
COMBINE CYCLE PLANT
ELEVATION

UNITED ENGINEERS & CONSTR. INC PHILA., PENNA.

-127

31.2



MEZZANINE FLOOR EL. 43'-0"

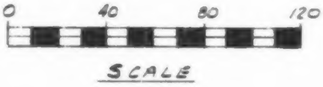


FIG. 1-128

635 M.W. FLUIDIZED BED BOILER
COMBINED CYCLE PLANT
MEZZANINE FLOOR PLAN

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PLATE VI

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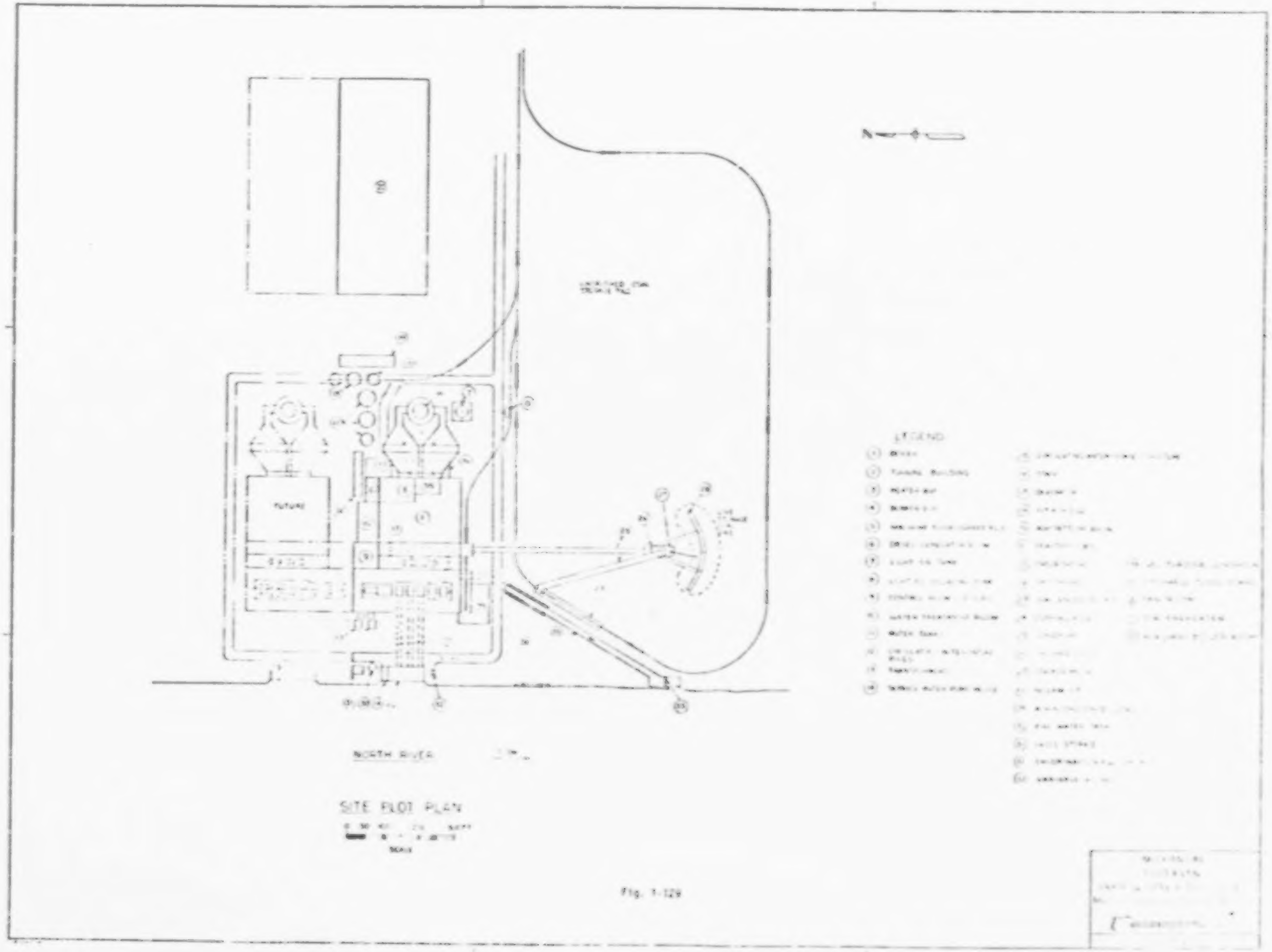

Fig. 1-129

335 /

335 3

III-2

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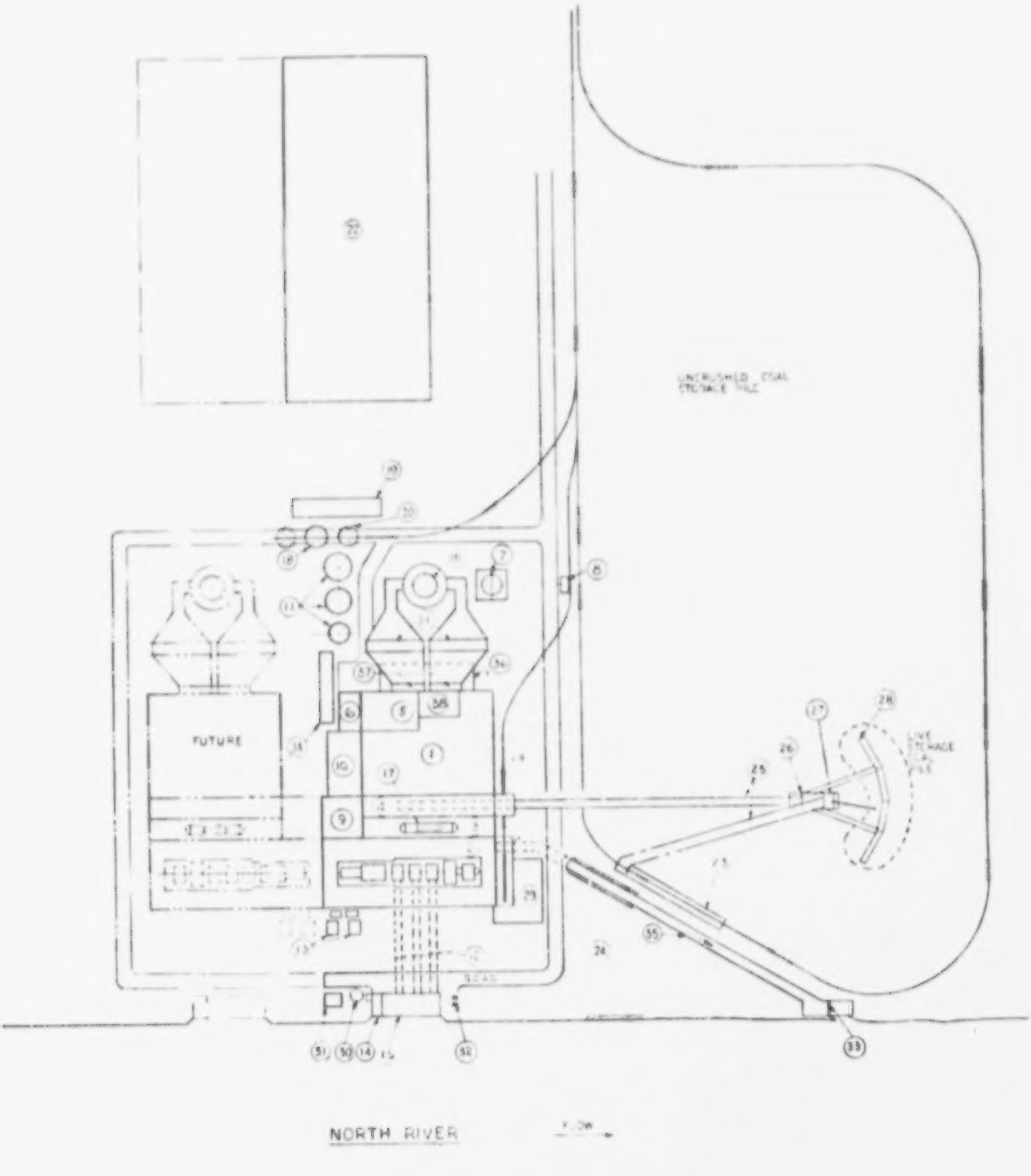


SITE PLOT PLAN

0 50 100 200 300 FT

SCALE

Fig. 1-129

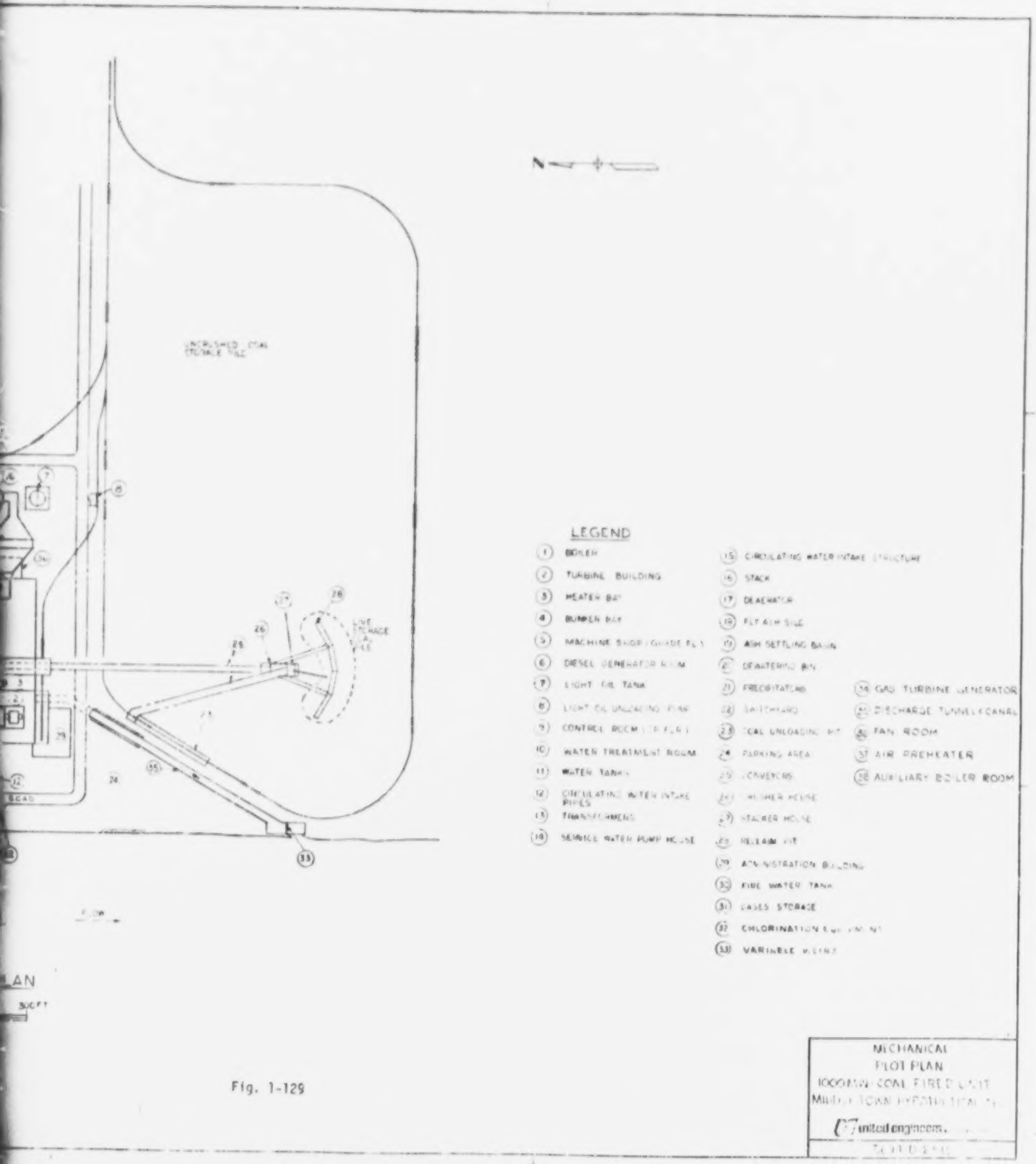
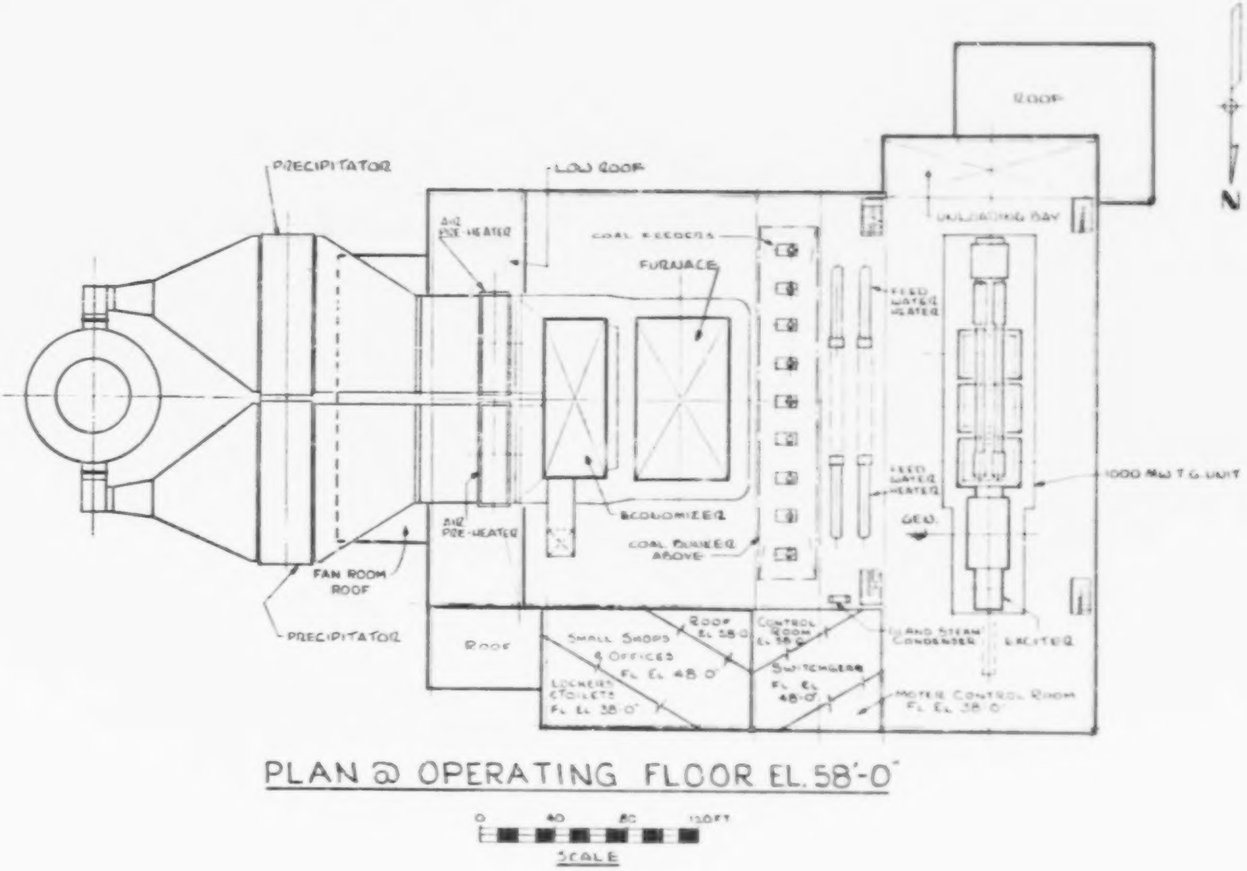

N
UNCRUSHED COAL STORAGE PILE
LIVE STORAGE COAL PILE
FLOW
LAN
300 FT
LEGEND
① BOILER
② TURBINE BUILDING
③ HEATER BAY
④ BUNKER BAY
⑤ MACHINE SHOP (GRADE FL.)
⑥ DIESEL GENERATOR ROOM
⑦ LIGHT OIL TANK
⑧ LIGHT OIL UNLOADING PUMP
⑨ CONTROL ROOM (OP. FL.)
⑩ WATER TREATMENT ROOM
⑪ WATER TANKS
⑫ CIRCULATING WATER INTAKE PIPES
⑬ TRANSFORMERS
⑭ SERVICE WATER PUMP HOUSE
⑮ CIRCULATING WATER INTAKE STRUCTURE
⑯ STACK
⑰ DEAERATOR
⑱ FLY ASH SILO
⑲ ASH SETTLING BASIN
⑳ DEWATERING BIN
㉑ PRECIPITATORS
㉒ SWITCHYARD
㉓ COAL UNLOADING PIT
㉔ PARKING AREA
㉕ CONVEYORS
㉖ CRUSHER HOUSE
㉗ STACKER HOUSE
㉘ RECLAIM PIT
㉙ ADMINISTRATION BUILDING
㉚ FIRE WATER TANK
㉛ GASES STORAGE
㉜ CHLORINATION EQUIPMENT
㉝ VARIABLE WEIRS
㉞ GAS TURBINE GENERATOR
㉟ DISCHARGE TUNNEL & CANAL
(36) FAN ROOM
(37) AIR PREHEATER
(38) AUXILIARY BOILER ROOM
MECHANICAL
PLOT PLAN
1000 MW COAL FIRED UNIT
MIDDLETOWN HYPOTHETICAL SITE
united engineers

Fig. 1-129

Fig. 1-130
MECHANICAL
GENERAL ARRGT-PLANS
1000 MW COAL FIRED UNIT
MIDDLETOWN HYPOTHETICAL SITE
united engineers



PLAN @ OPERATING FLOOR EL. 58'-0"

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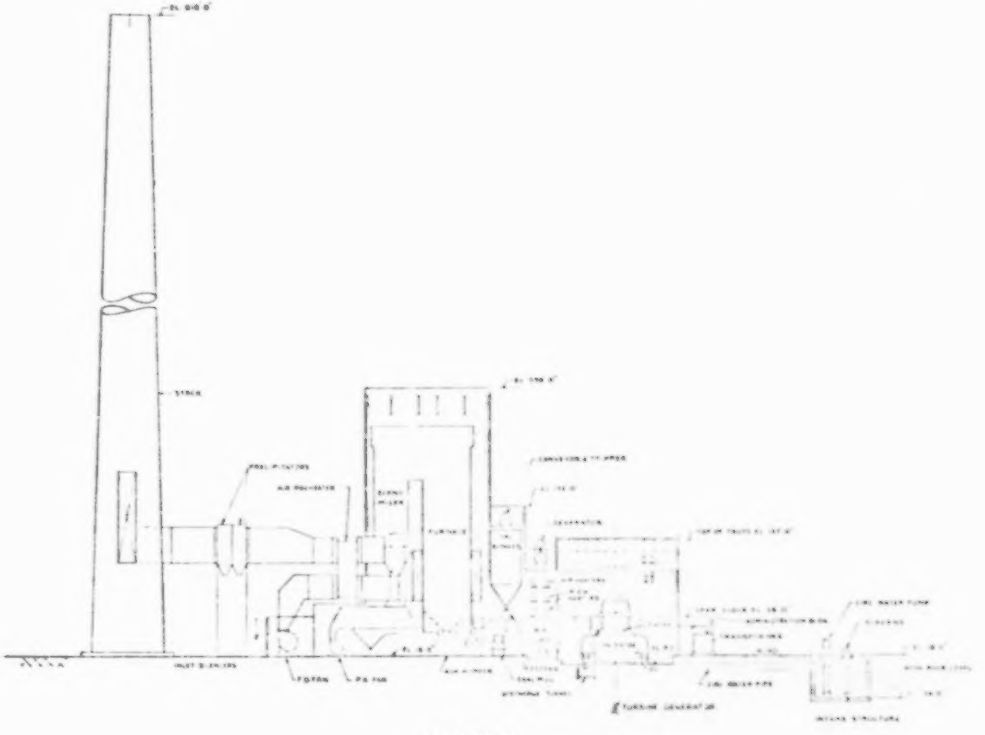


ELEVATION

SCALE

Fig. 1-132

MECHANICAL
GENERAL ARRGT ELEV
[illegible] MW COAL FIRED
[illegible]

3391

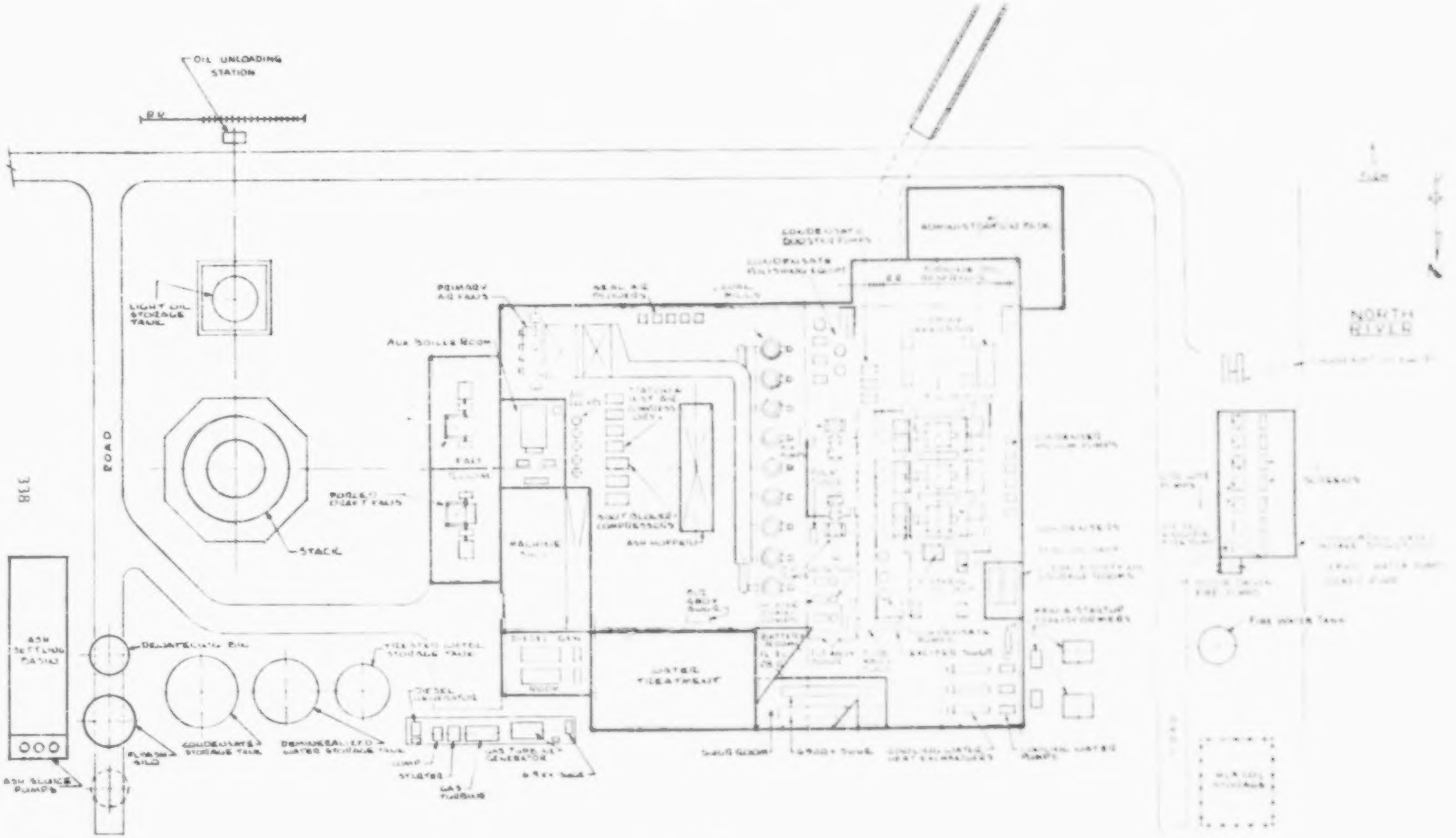


Fig. 1-131

PLAN @ GRADE EL. 18'-0"

0 40 80 120FT
SCALE

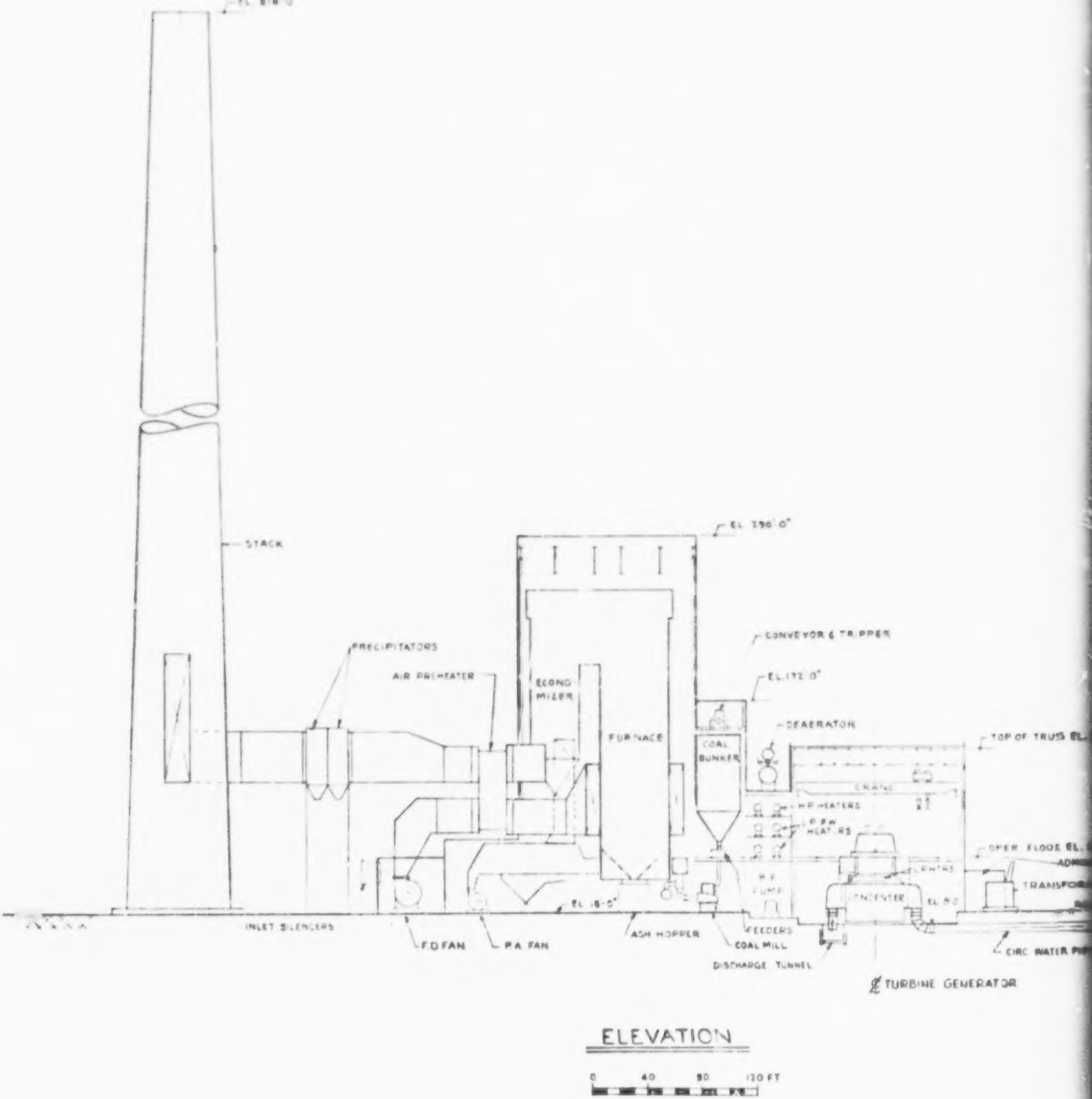

EL 818'0"
STACK
PRECIPITATORS
AIR PREHEATER
ECONO-MIZER
FURNACE
EL 290'0"
CONVEYOR & TRIPPER
EL 172'0"
COAL BUNKER
DEAERATOR
TOP OF TRUSS EL
CRANE
HP HEATERS
LP FW HEATERS
OPER. FLOOR EL
B.F. PUMP
CONDENSER
EL 8'0
TRANSFOR
INLET SILENCERS
F.D FAN
P.A FAN
EL 18'0"
ASH HOPPER
FEEDERS
COAL MILL
DISCHARGE TUNNEL
CIRC WATER PI
℄ TURBINE GENERATOR
ELEVATION
0
40
80
120 FT
SCALE

Fig. 1-122

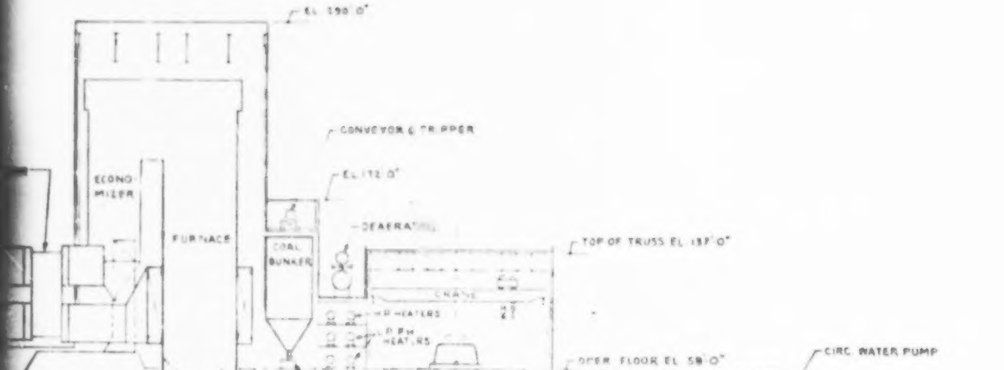
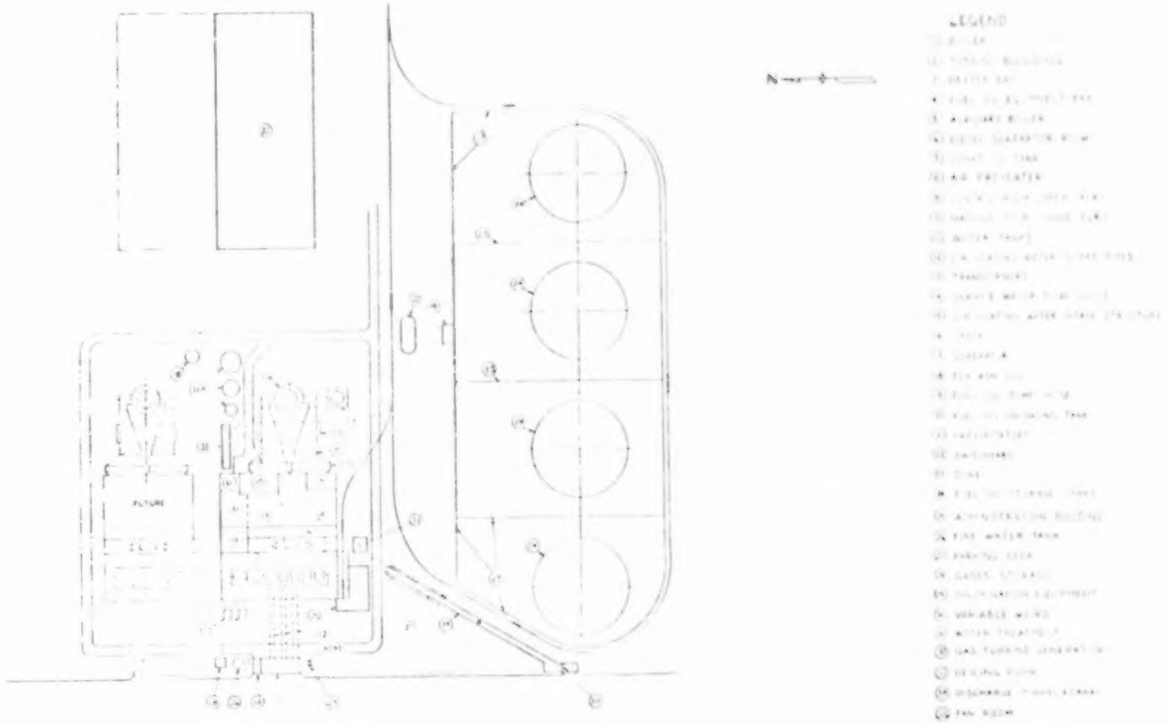


Fig. 1-132

MECHANICAL
GENERAL ARRGT ELEVATION
1000MW - COAL FIRED
MIDDLETOWN HYPOTHETICAL SITE
united engineers
9674-D-[illegible]



SITE PLOT PLAN

Fig. 1-133

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341.2

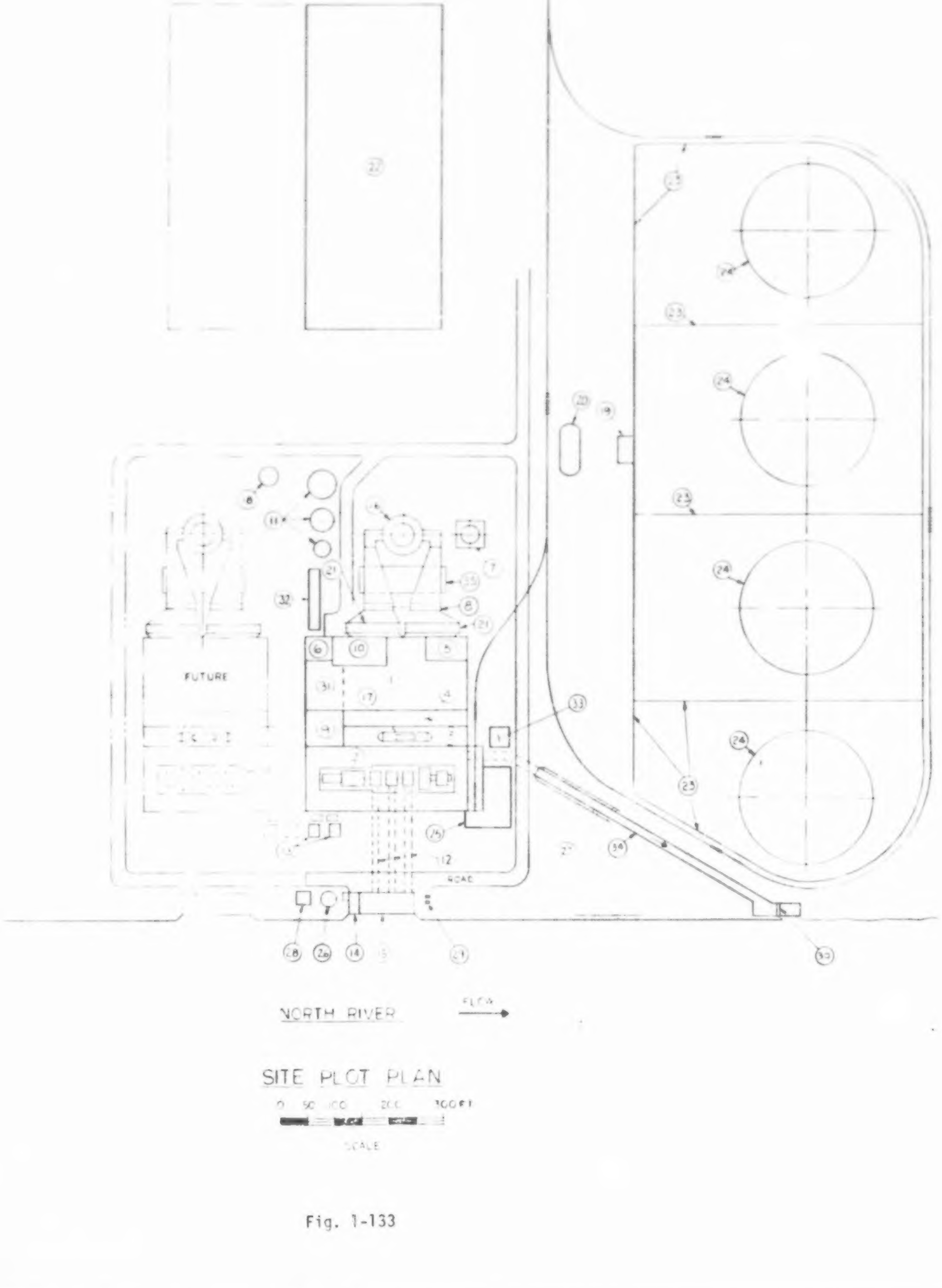

Fig. 1-133

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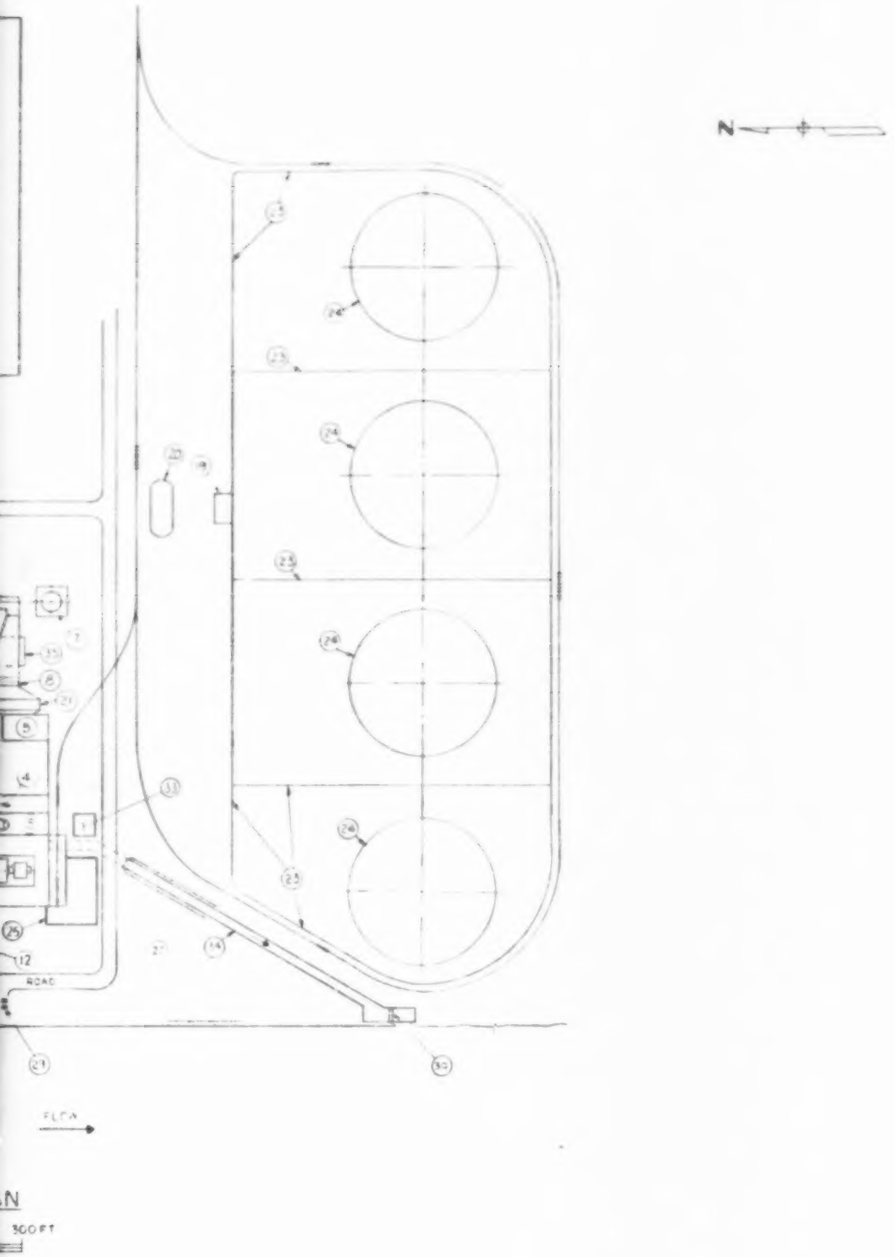


LEGEND

①	BOILER
②	TURBINE BUILDING
③	HEATER BAY
④	FUEL OIL EQUIPMENT BAY
⑤	AUXILIARY BOILER
⑥	DIESEL GENERATOR ROOM
⑦	LIGHT OIL TANK
⑧	AIR PREHEATER
⑨	CONTROL ROOM (OPER. FLR.)
⑩	MACHINE SHOP (GRADE FLR.)
⑪	WATER TANKS
⑫	CIRCULATING WATER INTAKE PIPES
⑬	TRANSFORMERS
⑭	SERVICE WATER PUMP HOUSE
⑮	CIRCULATING WATER INTAKE STRUCTURE
⑯	STACK
⑰	DEAERATOR
⑱	FLY ASH SILO
⑲	FUEL OIL PUMP HOUSE
⑳	FUEL OIL UNLOADING TANK
㉑	PRECIPITATORS
㉒	SWITCHYARD
㉓	DIKE
㉔	FUEL OIL STORAGE TANKS
㉕	ADMINISTRATION BUILDING
㉖	FIRE WATER TANK
㉗	PARKING AREA
㉘	GASES STORAGE
㉙	CHLORINATION EQUIPMENT
㉚	VARIABLE WEIRS
㉛	WATER TREATMENT
㉜	GAS TURBINE GENERATOR
㉝	DEICING PUMP
㉞	DISCHARGE TUNNEL & CANAL
㉟	FAN ROOM

3.8.11

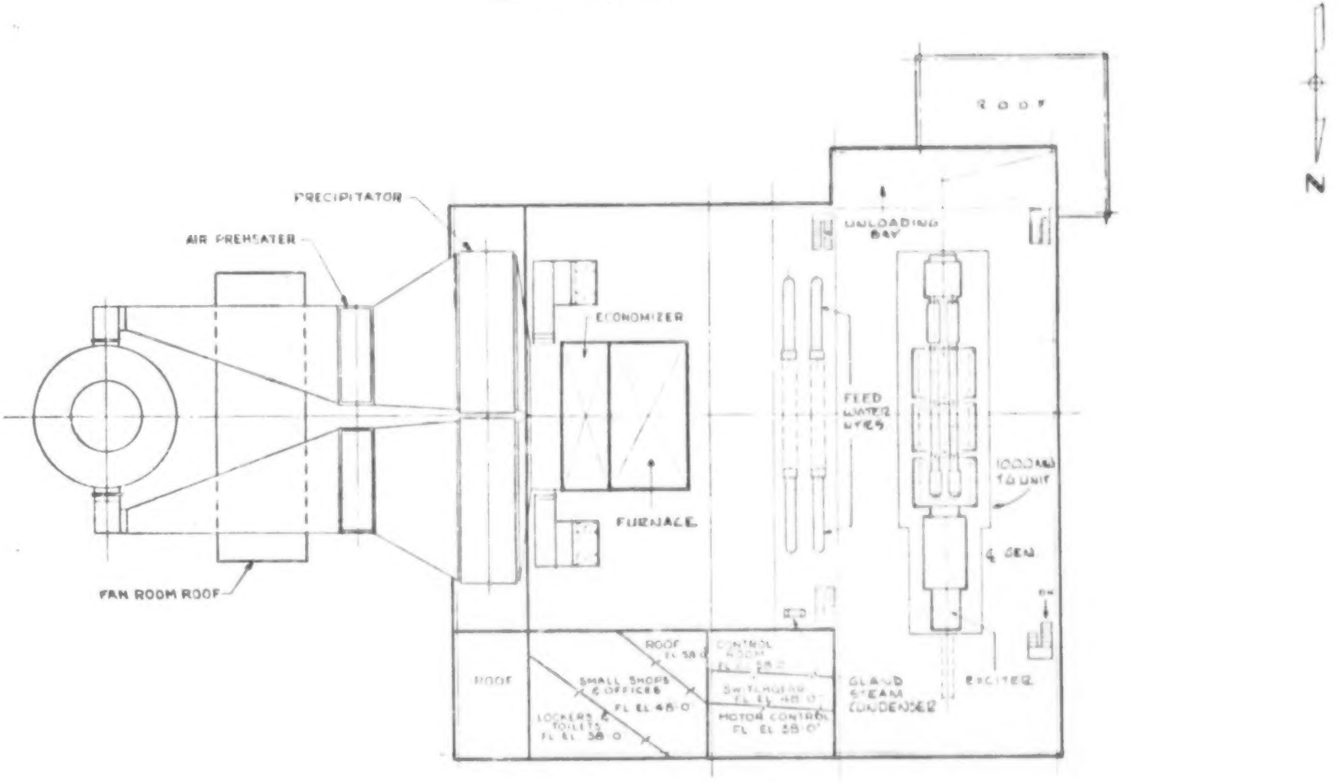
MECHANICAL
PLOT PLAN OF SITE
1000 MW- OIL FIRED UNIT
MIDDLETOWN HYPOTHETICAL SITE

united engineers & constructors

9674-D-2521

Fig. 1-134

MECHANICAL
GENERAL ARRGT PLANS
1000 MW OIL FIRED
MIDDLETOWN HYPOTHETICAL SITE
united engineers.



PLAN @ OPERATING FLOOR EL. 58'-0"

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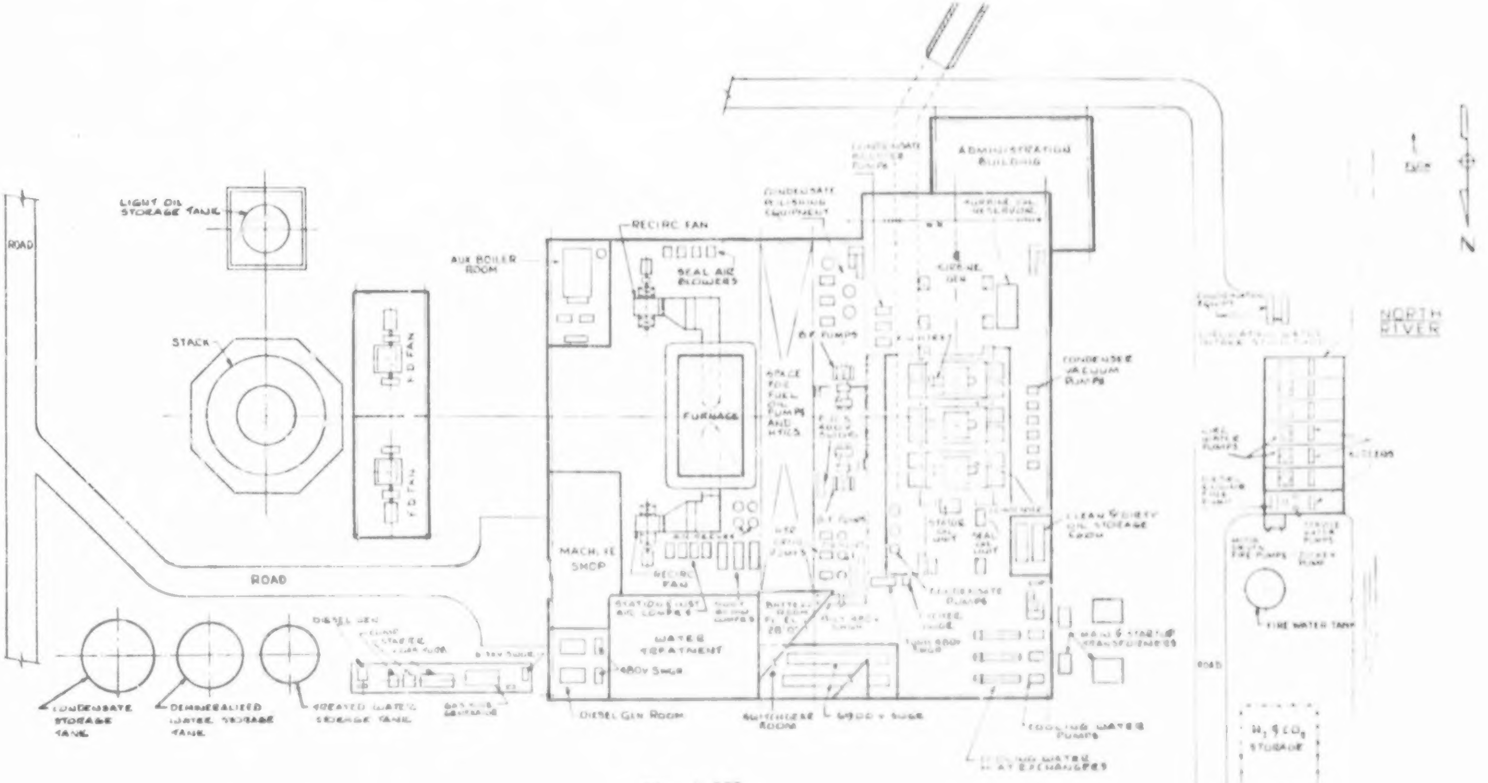

Fig. 1-135

PLAN@GRADE EL 18'-0"

0 40 80 120 FT

SCALE

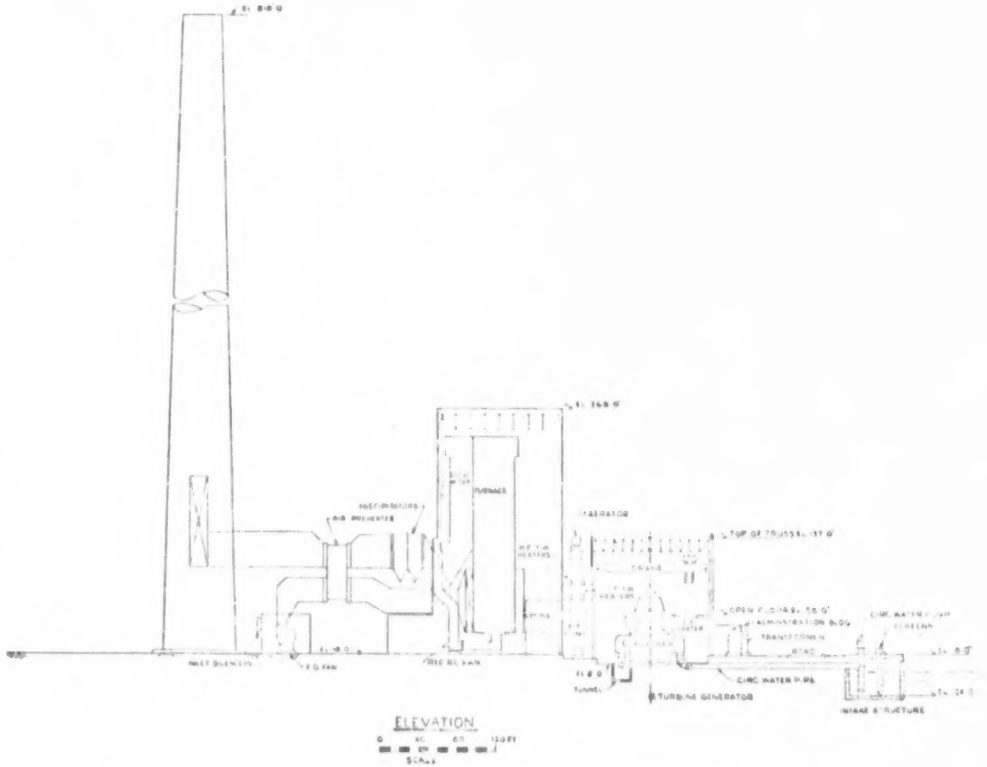

Fig. 1-136

MECHANICAL
GENERAL ARR'GT ELEVATION
1000MW [illegible] FIRED
[illegible]

united engineers

345.1

345.2

345.3

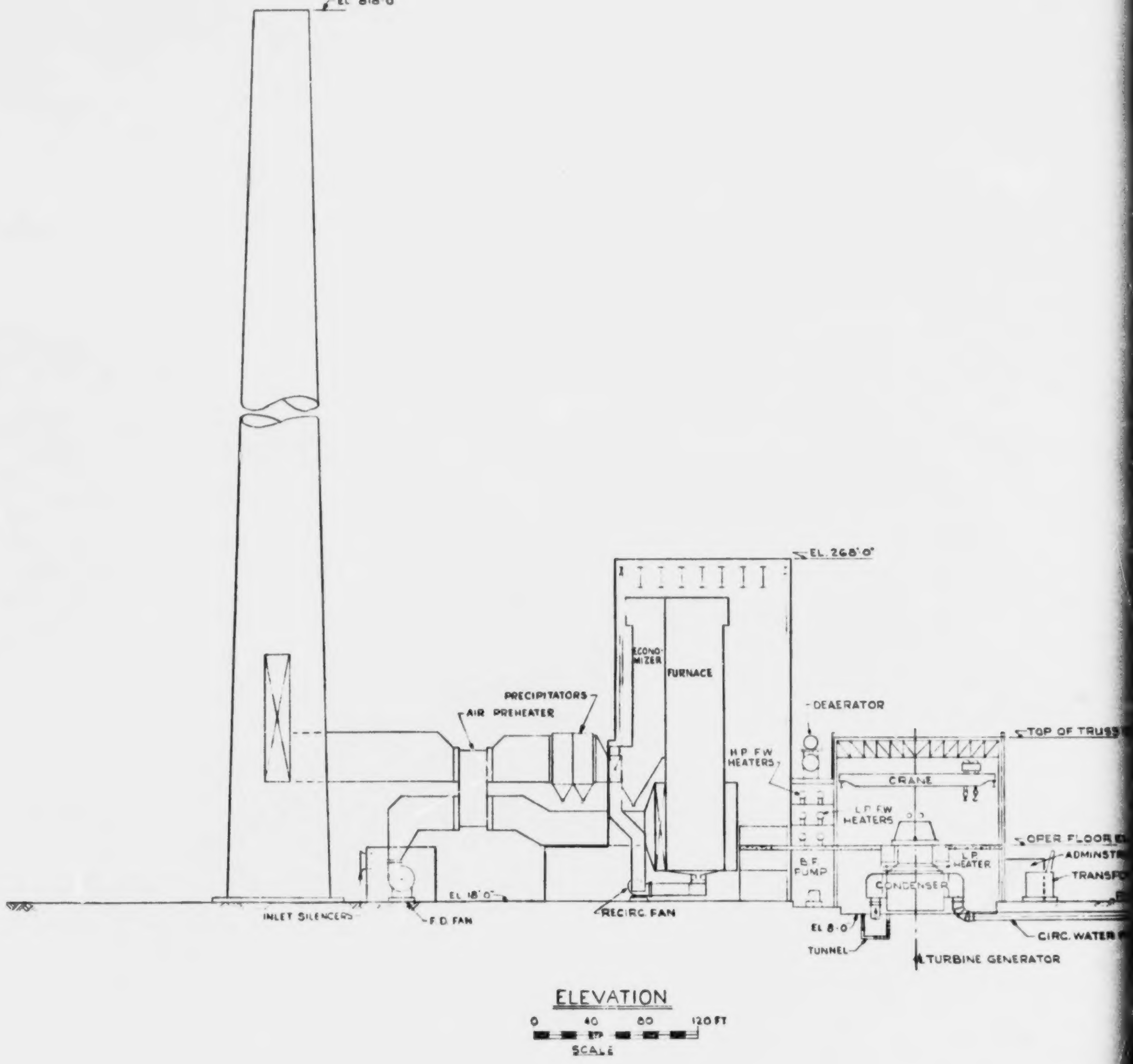
EL 818'0"
EL. 268'0"
ECONO-MIZER
FURNACE
PRECIPITATORS
AIR PREHEATER
DEAERATOR
H P F W HEATERS
TOP OF TRUSS
CRANE
L P F W HEATERS
OPER FLOOR EL
ADMINSTR
TRANSF
B F PUMP
L P HEATER
CONDENSER
EL 18'0"
INLET SILENCERS
F D FAN
RECIRC FAN
EL 8-0
TUNNEL
CIRC. WATER
℄ TURBINE GENERATOR
ELEVATION
0
40
80
120 FT
SCALE

Fig. 1-136

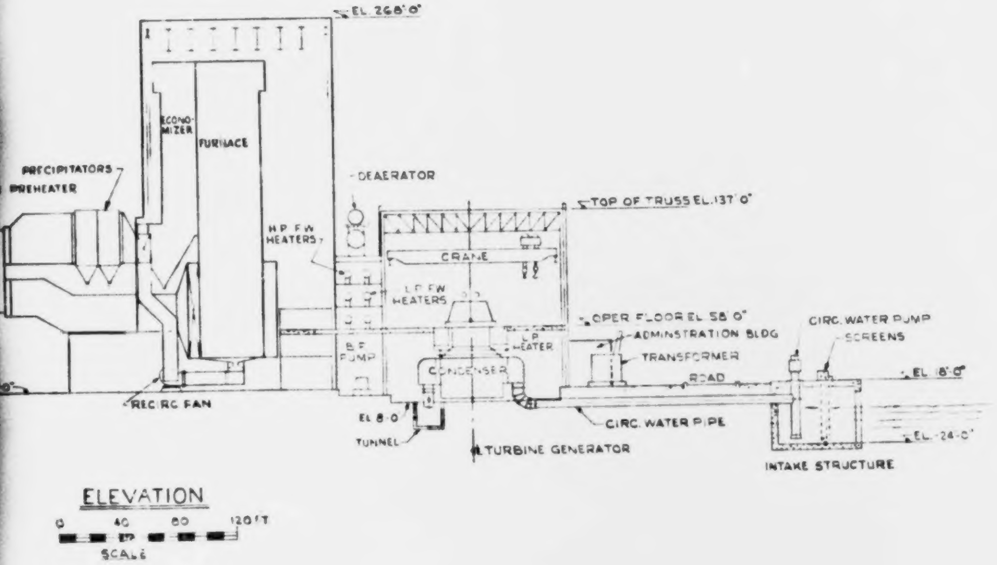

Fig. 1-136

MECHANICAL
GENERAL ARRGT. ELEVATION
1000 MW OIL FIRED
MIDDLETOWN HYPOTHETICAL SITE
united engineers
1 D 2526

Component	
Air outlet transition	0.75
Piping between transition and boiler	0.20
Control valves in open position	0.10
Internal air passages to plenum	0.05
Distribution plate	1.00
Bed	3.10
Internal gas passage to inlet of first stage collector	0.10
First-stage collector	0.48
Second-stage collector (4 units)	0.77
Piping from second-stage separator to transition	0.20
Hot gas inlet transition	0.75
Total	7.50

An alternative arrangement for the second-stage collector is being considered wherein two tornado collectors in parallel are used in place of four. The pressure loss for this configuration is 1.38%, giving a total system pressure drop of 8.11%.

The full load heat rates for these sets of conditions are as follows:

BACK PRESSURE IN. Hg	COMBUSTION EFFICIENCY-%	PRESSURE DROP-%	PLANT HEAT RATE-BTU/kwh
3	89.6	3	8974
1-1/2	88.6	7.5	8892
1-1/2	88.6	8.1	8897
3	88.6	3	9075

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Recommendations

Pressurized fluidized bed power plants show greater pollution abatement, lower costs, and greater development potential than atmospheric pressure fluidized bed or conventional p.f. plants. Based on the projected market for coal and the technical assessment, we recommend that:

- Emphasis be placed on the development of pressurized fluidized bed power plants
- Alternate and advanced concept evaluation be conducted to assess potential
- A 10-30 MW pressurized fluidized bed boiler development plant be designed and constructed
- Laboratory research and development on pressurized fluidized bed combustion continue in key areas: sulfur removal, NO_x reduction, stone regeneration, sulfur recovery, materials evaluation, advanced concepts, particulate removal, and turbine blade erosion.

Atmospheric Pressure Fluidized Bed Boiler Design

Specifications

Specifications for the power plant and the fluidized bed boiler were prepared by Westinghouse for use by Foster Wheeler in making a preliminary design of the atmospheric boiler.

Plant. The specifications for the power plant for which the atmospheric pressure fluidized bed boiler was to be designed were as follows:

- Capacity - 300 MW
- Type - intermediate load
- Turn-down requirements - 4:1
- Transient requirements - 5% per min.
- Geographical location - Eastern United States
- Pollution control
  - Particulate emissions - 0.01 gr/SCF
  - Sulfur removal - 90%
  - Nitrogen oxide emissions - <0.7 lb/10^6 BTU

Boiler. The specifications for the atmospheric pressure fluidized bed boiler were as follows:

- Fuel - Ohio Pittsburgh No. 8 Seam Coal

Proximate analysis (wt %)

Volatile matter	39.5
Fixed carbon	48.7
Ash	8.5
Moisture (as mined)	3.3
	100.0

Sulfur content - 4.3%

Particle size - 1/4 in x 0

Complete data on coal is given on page 167.

- Sorbent - 1359 Limestone

  Particle size - -1/4 in with minimum fines
  Feed ratio - 6 x stoichiometric
  Complete data on sorbent is given on page 170

- Bed Design Parameters

  Primary beds
    Temperature - 1600°F design pt.
      1700°F max.
      1400°F min.
    Size distribution of material in bed
      Max. 5000 μ
      Min. 1000 μ
      Mean 2500 μ
    Superficial velocity - 10 to 15 ft/sec
    Excess air - 10%
    Combustion efficiency - 87% design pt.
    Density - 45 lb/ft^3
    Expanded height - 2-1/2 to 3 ft
    Elutriated material
      Ash - 100%
      Coal - 13%
      Sorbent - 3% of recirculated limestone
      Size distribution - see page 127
    Gas-side heat transfer coefficients
      In bed - 50 BTU/ft^2 hr - °F
      Above bed - 40 BTU/ft^2 hr - °F

- Carbon Burn-up Cell

  Temperature - 1900°F design
    Range - 1800°F to 2000°F
    Size distribution of material in bed
      Max. 5000 μ
      Min. 1000 μ
      Mean 2500 μ

Superficial velocity - 10 to 15 ft/sec
Excess air - at least 50%
Combustion efficiency - 85% design pt.
Density - 45 lb/ft^3
Expanded height - 2-1/2 to 3 ft
Recovery of coal elutriated from primary beds - >90%
Elutriated material
    Ash - 100%
    Coal - 15%
    Sorbent - 100% of sorbent in with recovered coal plus 3% of recirculated limestone.
    Size distribution - see page 127
Gas-side heat transfer coefficients
    Same as for primary beds

Stack gas temperature - 275°F
Air preheat temperature - to be determined

Cycle Selection

Consideration was given to a subcritical cycle (2400 psig/1000°F/1000°F) and a supercritical cycle (3500 psig/1000°F/1000°F). The subcritical cycle was selected for this intermediate load application because of its advantages in reliability, availability, and delivery and start-up. Also, in practice the supercritical cycle plants have not demonstrated a significant advantage in unit heat rate. Steam temperatures currently in use in pulverized coal units were selected, although the possibility of reduced boiler tube corrosion in fluidized beds indicates that this is a conservative position.

A limited study was made to determine the effect of increasing the steam temperature on turbine power and plant net heat rate. A condenser pressure of 3 in. Hg. was used in these calculations.

STEAM PRESSURE PSIA	SUPERHEAT TEMP. °F	REHEAT TEMP. °F	STEAM FLOW %	TURBINE POWER %	PLANT HEAT RATE %
2400	1000	1000	100	100	100
3300	1100	1100	100	110.7	95
4500	1200	1100	100	118.0	92.1

This study indicates that the use of a fluidized bed boiler may provide a significant growth potential for conventional Rankine cycle systems.

The once-through system was selected over the natural circulation system for the subcritical cycle because it permits the use of horizontal boiler tubes in the fluidized beds.

The steam system used in this boiler design is that from the Hammond No. 4 unit of Georgia Power and Light (see following matrix). This system has seven regenerative feedwater heaters and its turbine heat rate versus net load is plotted in Figure 1-137.

Boiler Design

Alternative Concepts. Three basic concepts were considered for the preliminary design of the atmospheric pressure fluidized bed boiler:

- Vertically stacked beds with four modules in a contiguous square array
- Vertically stacked beds in four separate modules in an in-line arrangement
- Horizontal tandem arrangement with four modules in parallel.

An early decision to use a modular arrangement with each bed having a single function precluded modules having a single bed and a single module boiler. The modular arrangement is necessitated by turn-down requirements and facilities start-up.

STEAM CYCLE CONDITIONS

300 MW ATM FLUIDIZED BED BOILER

Primary steam flow	$1.9\text{x}10^6$ lb/hr
Steam pressure, superheater out	2400 psig
Reheater in	601 psig
Reheater out	581 psig
Steam temperature, superheater out	1000°F
Reheater in	650°F
Reheater out	1000°F
Steam flow (reheater)	$1.6\text{x}10^6$ lb/hr
Boiler feedwater temperature	480°F

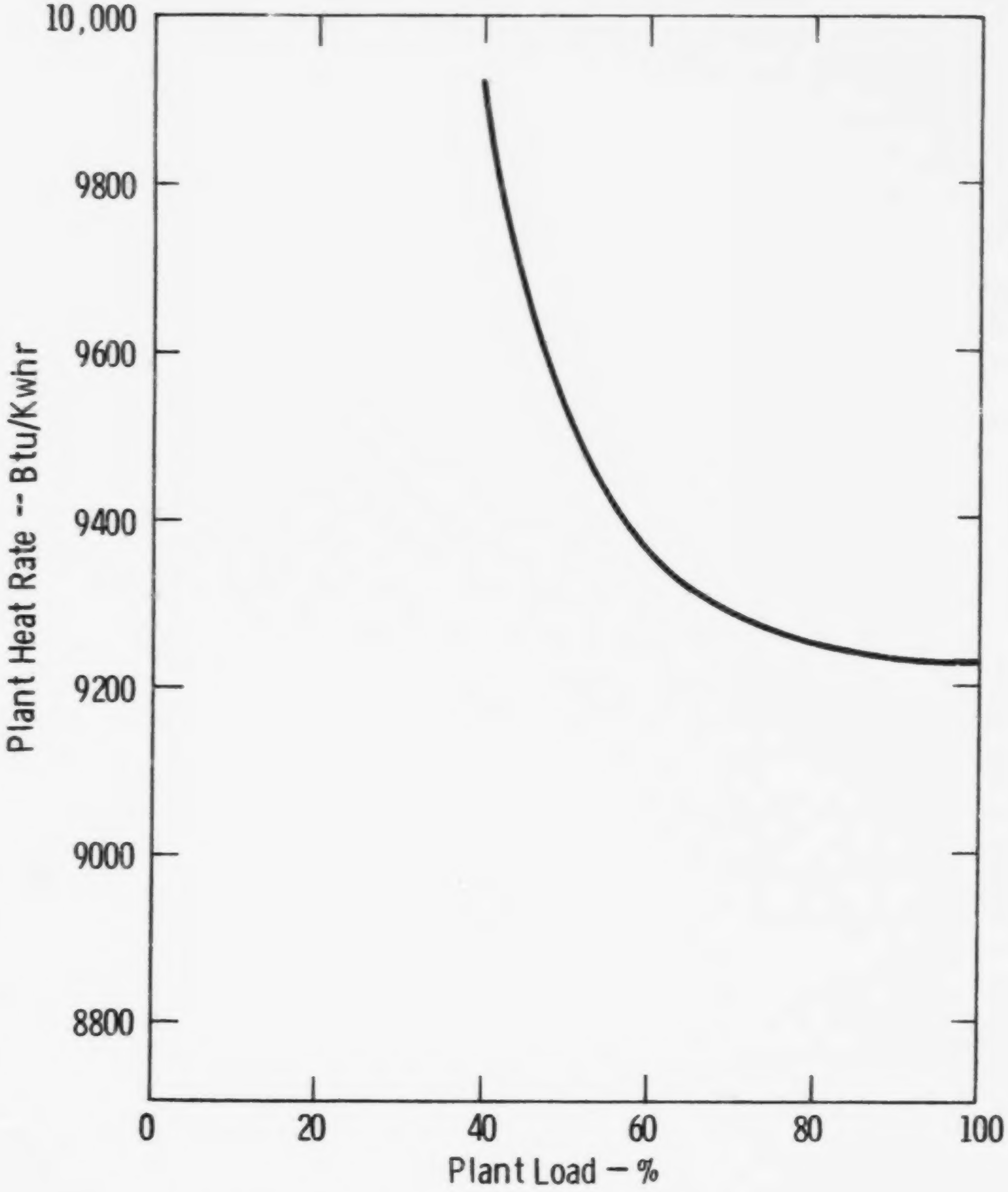


Fig. 1-137 — Part load performance of Hammond #4 unit 2400/1000/1000/3 In.

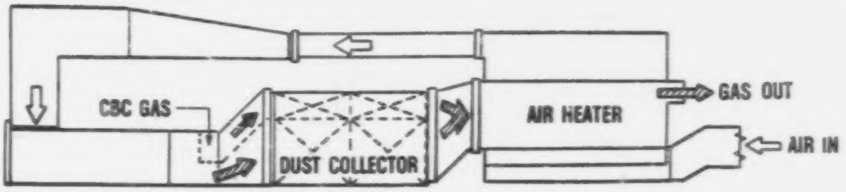
Preliminary Design of Preferred Concept. The in-line arrangement of four separate modules of vertically stacked beds was selected as the preferred concept (see Figure 1-138). The vertical beds were chosen over the horizontal beds to simplify the steam piping, the distribution of air to the beds, and the removal of products of combustion from the beds. The in-line arrangement was chosen over the square array because there were more disadvantages than advantages in sharing common walls, and symmetry of auxiliary equipment was lacking so that ductwork and piping would have been complicated.

The preliminary design configuration consisted of four modules, each containing six beds, including a carbon burn-up cell. Two of the beds are assigned to evaporating functions, two to superheating, and one to reheating. The carbon burn-up cell bed has no immersed tubes. The water walls used for enclosing the beds are evaporating tubes. Preheating of the feedwater is done in separate convection passes located in the overhead above each bed. The advantages of this over an external economizer are:

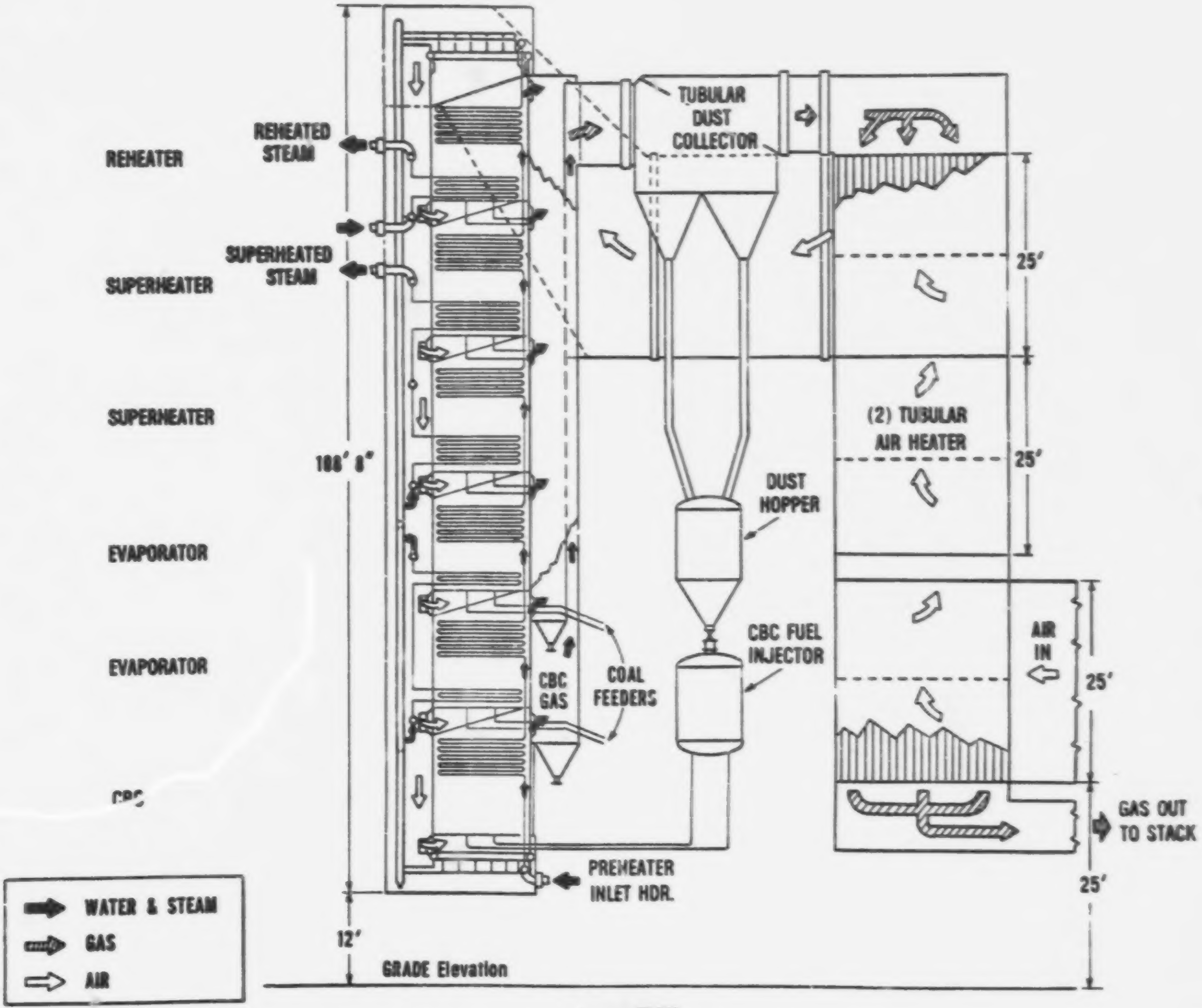
- The gases are cooled within the boiler enclosure, thus eliminating the need for high-temperature ducts leading to an economizer
- The convection surface acts as a particle screen and helps in the detrainment of the larger particles elutriated from the beds
- It makes more effective use of the boiler cross sectional area.

The dimensions of the beds in the four module boilers are 13 ft x 12 ft x 2-1/2 ft, and the module height is about 100 ft. Since the overall dimensions exceed the limits for land transport, field erection will be necessary. However, the maximum degree of shop fabrication of subassemblies will be employed.

Horizontal tube bundles are used in all beds and for the convection passes above the beds. All bundles are arranged in a rectangular



HALF PLAN VIEW



ELEVATION

ATMOSPHERIC–PRESSURE FLUIDIZED BED STEAM GENERATOR

FOUR (4) REQUIRED

Fig. 1-138

pattern, and the tubes are on a square or rectangular array. The submerged tubes have both vertical and horizontal pitches of 4 in. Two-inch O.D. tubes are used in the reheater bundle and 1-1/2-inch O.D. tubes in the superheater and evaporator. In the preheater bundles, the first two loops have a vertical pitch of 4 in and a horizontal pitch of 8 in. The remainder of the bundle is on a 4 in x 4 in square array. The preheater tubes are 2 in O.D.

The water walls which form the enclosure for the beds are of welded fin configuration. They are constructed of 1-3/4-in tubes on 2-in centers with a 1/4-in fin. This spacing is the minimum practical with this type of construction.

The water-steam circuitry is shown on an isometric flow diagram in Figure 1-139. Water enters the feedwater inlet header and then proceeds in series through the preheater bundles which are located in the convection passes above each bed. From the feedwater outlet header at the top of the module, the water goes to the first evaporation bundle in the bed above the carbon burn-up cell. This is in series with the second evaporation bundle in the next bed above. Evaporation is completed in the water wall enclosure. The four walls are connected in series.

The evaporated water then goes through the two bundles of superheater tubes in series. A mixing header is located between the two superheater bundles. The superheated steam leaves the module through a single-ended header and goes to the high-pressure steam turbine. Reheat steam enters the module through a header at the inlet end of the reheat bundle and leaves through a header at the exit end ot the bundle. Both headers are single-ended.

The auxiliary equipment for the atmospheric pressure fluidized bed steam generator includes an air preheater, a coal handling and storage system, a combined coal drying and feeding system, a limestone handling and feeding system, a limestone regeneration system, and fans and particulate removal equipment.

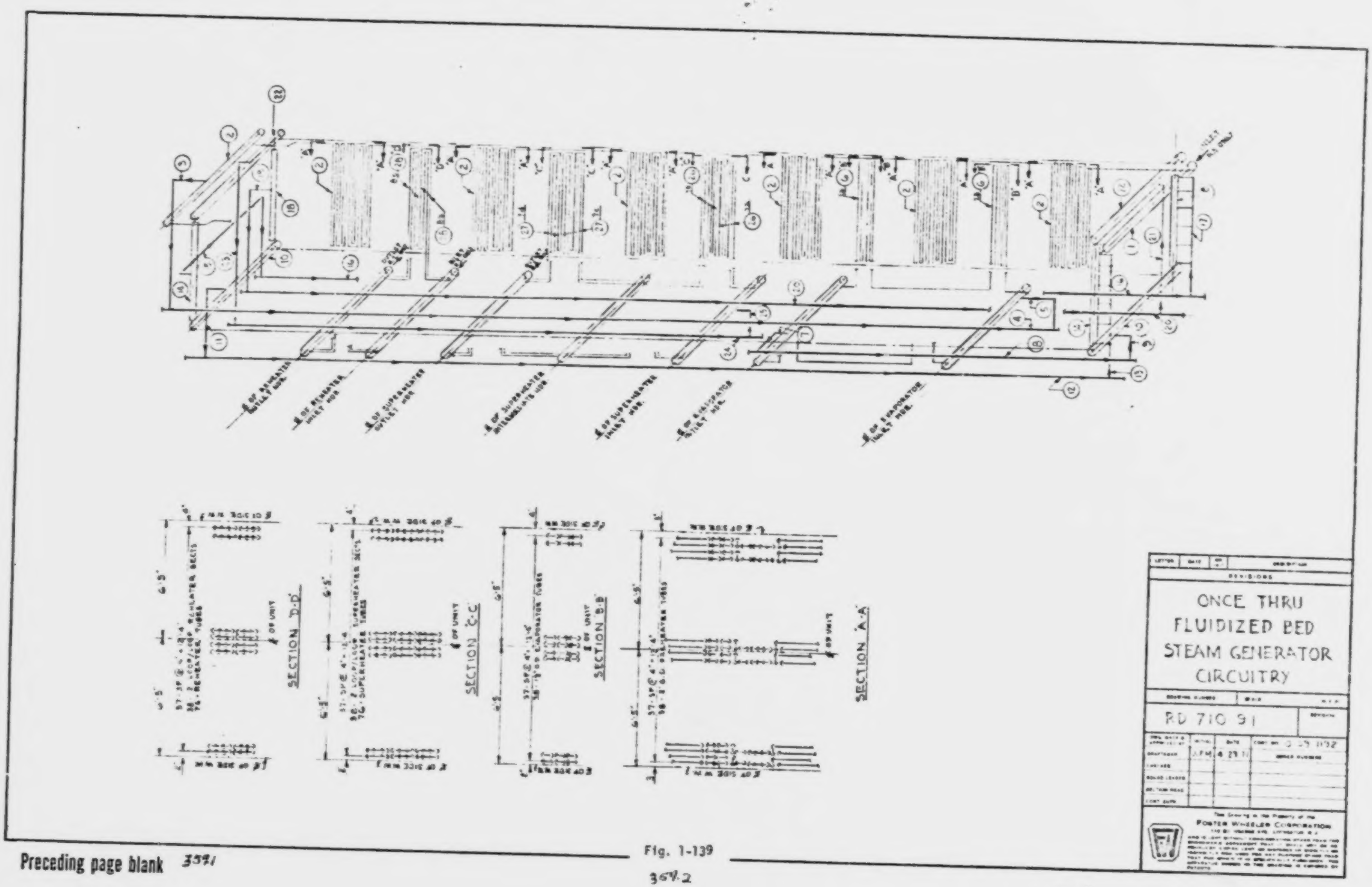

℄ OF REHEATER OUTLET HDR.
℄ OF REHEATER INLET HDR.
℄ OF SUPERHEATER OUTLET HDR.
℄ OF SUPERHEATER INTERMEDIATE HDR.
℄ OF SUPERHEATER INLET HDR.
℄ OF EVAPORATOR OUTLET HDR.
℄ OF EVAPORATOR INLET HDR.
INLET
SECTION 'D-D'
SECTION 'C-C'
SECTION 'B-B'
SECTION 'A-A'
℄ OF UNIT
6'-5"
REVISIONS
ONCE THRU
FLUIDIZED BED
STEAM GENERATOR
CIRCUITRY
RD 710 91
DRAFTSMAN
AFM
4 29 71
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Fig. 1-139

359.2

359.3

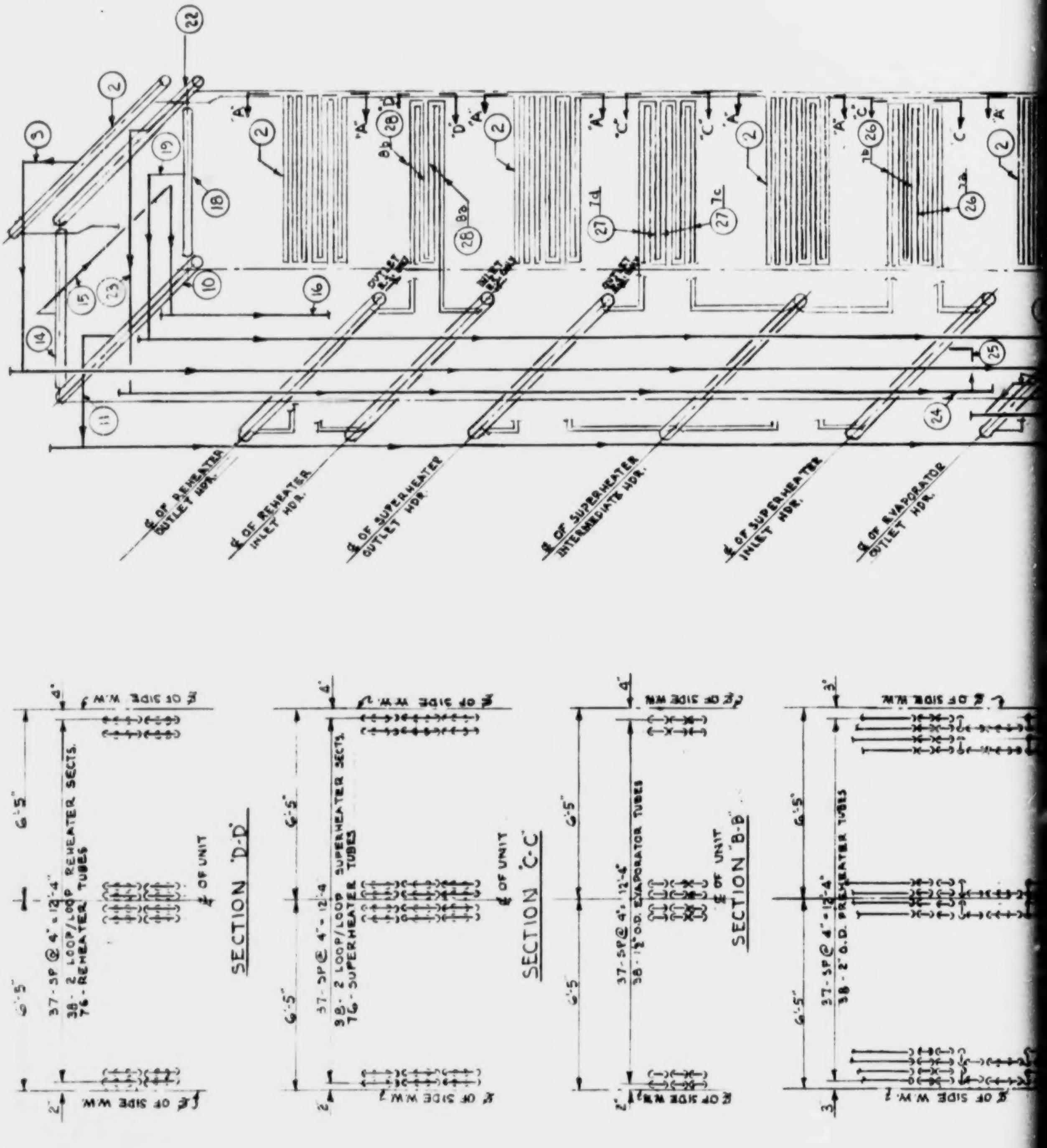
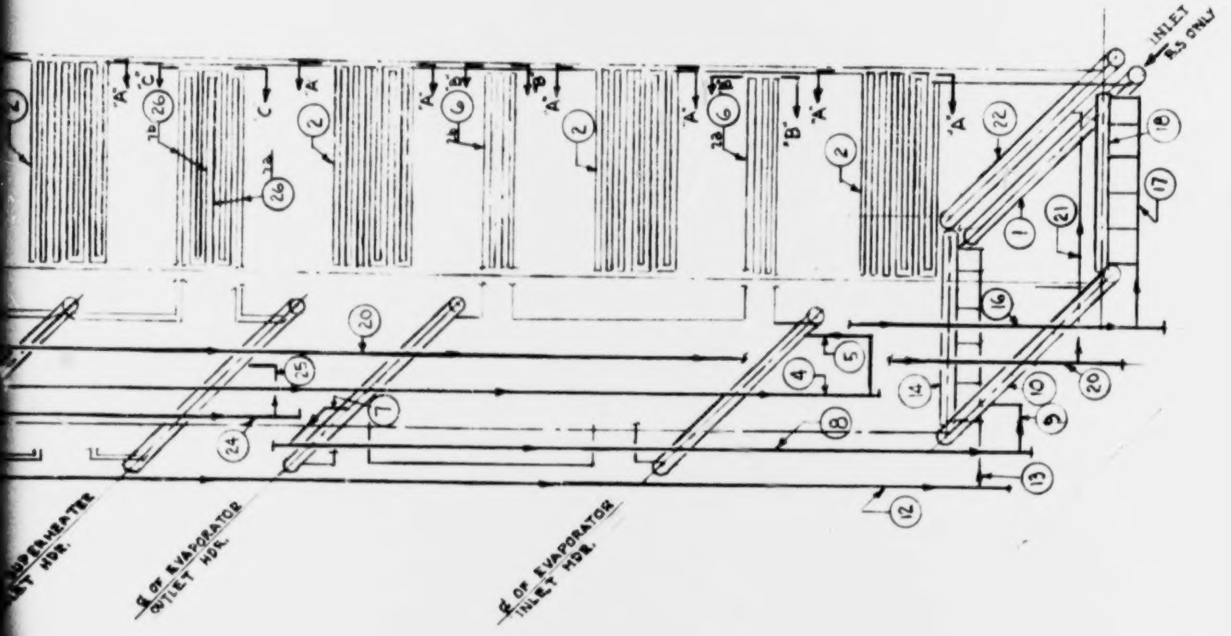
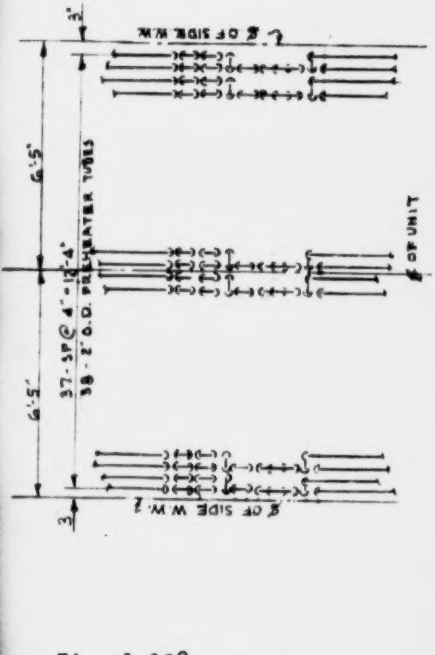

℄ OF REHEATER OUTLET HDR.
℄ OF REHEATER INLET HDR.
℄ OF SUPERHEATER OUTLET HDR.
℄ OF SUPERHEATER INTERMEDIATE HDR.
℄ OF SUPERHEATER INLET HDR.
℄ OF EVAPORATOR OUTLET HDR.
SECTION 'D-D'
37-SP @ 4" = 12'-4"
38 - 2 LOOP/LOOP REHEATER SECTS.
76 - REHEATER TUBES
SECTION 'C-C'
37-SP @ 4" = 12'-4"
38 - 2 LOOP/LOOP SUPERHEATER SECTS.
76 - SUPERHEATER TUBES
SECTION 'B-B'
37-SP @ 4" = 12'-4"
38 - 1½" O.D. EVAPORATOR TUBES
37-SP @ 4" = 12'-4"
38 - 2" O.D. PREHEATER TUBES
℄ OF UNIT
℄ OF SIDE W.W.
6'-5"
4"
2"
3"

Fig. 1-139





SECTION 'A-A'

LETTER	DATE	DR. INIT.	DESCRIPTION
REVISIONS			

ONCE THRU
FLUIDIZED BED
STEAM GENERATOR
CIRCUITRY

DRAWING NUMBER	SCALE: = 1'-0"
RD 710-91	REVISION

DWG. MADE & APPROVED BY	INITIAL	DATE	CONT. NO. 0-59-1192
DRAFTSMAN	J.P.M.	4-29-71	ORDER NUMBERS
CHECKER			
SQUAD LEADER			
SECTION HEAD			
CONT. SUPV.			

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Fig. 1-139

Air from the air preheater passes through two external ducts to a common internal duct which runs the full height of the boiler module. From this common duct, air is metered by dampers into plenum chambers below the fluid bed distributor plates. A separate air stream which by-passes the preheater goes directly from the forced draft fan to the carbon burn-up cell.

An internal common duct collects the flue gas with carbon-rich ash from the five primary fluidized beds and carries it to the primary cyclone separator. A separate duct located within the main flue gas duct carries the flue gas from the carbon burn-up cell to another dust collector.

Coal is pneumatically transported through sixteen separate feed lines to each bed as a dilute phase of dried coal in 250°F air. The coal feed lines enter the gas duct from either side of the module, pass through the flue gas exit openings in the water wall, then up into the plenum chambers and through the grid plates. The coal is injected into the bed in a horizontal direction in a volume free of tube bundles. One coal injector is provided for each ten square feet of fluid bed surface. The recycle ash feed lines are brought into the module at a point below the carbon burn-up cell.

Sorbent feeding into the fluidized beds falls into two categories; make-up feed and recycled sorbent feed. The make-up sorbent is fed into the beds along with the coal. Each bed is provided with one injection point for the regenerated sorbent and one withdrawal point for the spent sorbent.

The boiler modules are not bottom supported, but hang from structural steel.

Regeneration - Sulfur Recovery System

A "direct reduction" process was selected for regeneration of the sulfated limestone. The calcium sulfate is converted to CO by reaction with a reducing gas

$$CaSO_4 + \begin{matrix} CO \\ H_2 \end{matrix} \rightarrow CO + SO_2 + \begin{matrix} CO_2 \\ H_2O \end{matrix} \tag{1-40}$$

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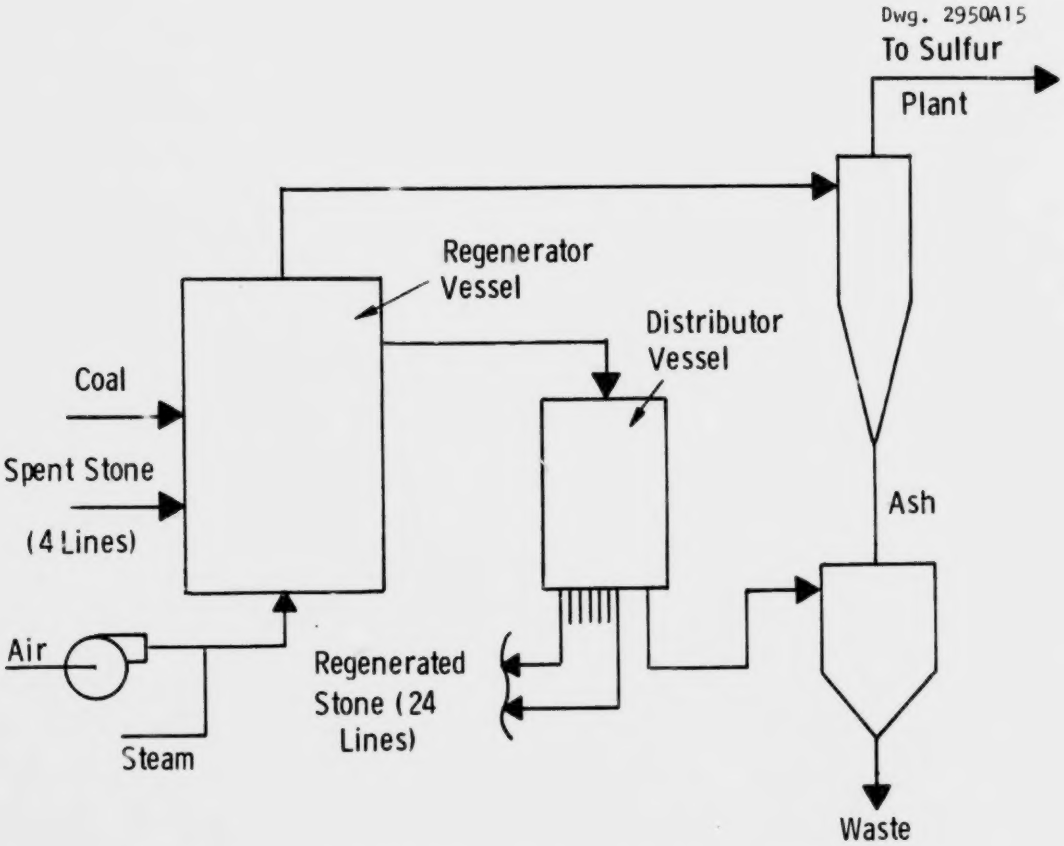
A flow diagram and material balance are given in Figure 1-140 and the following matrix for a 300 MW plant. Figure 1-141 is a schematic plant layout showing the boiler-regenerator interconnections.

MATERIAL BALANCE FOR 300 MW ATMOSPHERIC PRESSURE REGENERATION SYSTEM

Spent stone	56.6 tph = 113,000 lb/hr 37.8 wt % $CaSO_4$ 1500°F
Regenerated stone	43.5 tph 1.33 wt % $CaSO_4$ 2000°F
Waste stone	~9000 lb/hr
Air	2800 lb mole/hr
Coal	5.6 tph
Ash	0.48 tph
Sulfur gas	3500 lb mole/hr 2000°F
SO_2	6.3% (mole %)
H_2O	13.3%
CO_2	18.9%
N_2	61.1%
CO	0.3%
H_2	0.1%
Adiabatic heat	9 MMBtu/hr

Spent stone discharges from each boiler bed into a hold vessel, one hold vessel per boiler module. The stone is then pneumatically transported to the regenerator vessel, using air as the carrier gas.

The regenerator vessel is a fluid bed reactor operating at 1 to 2 atm. and 2000°F. Coal is charged into the reactor bed. An air-stream mixture fluidizes the reactor bed and combusts the coal to provide reducing gas for the $CaSO_4$ reaction.



Spent Stone: 56.6 tph = 113,000 lb/hr, 37.8 wt % $CaSO_4$, 1500°F
Regenerated Stone 43.5 tph = 87,000 lb/hr, 1.33 wt % $CaSO_4$, 2000°F
Waste Stone: ~ 9000 lb/hr
Air: 2800 lb mole/hr
Coal: 5.6 tph
Ash: 0.48 tps
Sulfur Gas: 3500 lb mole/hr 2000°F

SO_2: 6.3% (mole %)	N_2: 61.1%
H_2O: 13.3%	CO: 0.3%
CO_2: 18.9%	H_2: 0.1%

Adiabatic Heat: −9 MMBTU/hr

Fig. 1-140—Atmospheric pressure regeneration flow diagram

Dwg. 2950A13

Atmospheric-Pressure 300 MW Boiler System

1. Boiler Module..........16' x 16' ea
2. Regenerator Vessel.....See Sketch
3. Distributor Vessel.....12' Ø x 12' ht.
4. Hold Drums.............5' Ø x 5' ht.
5. Cyclone ....
6. Waste Stone Hopper.....8' Ø x 12' ht.
7. Air Blower.............

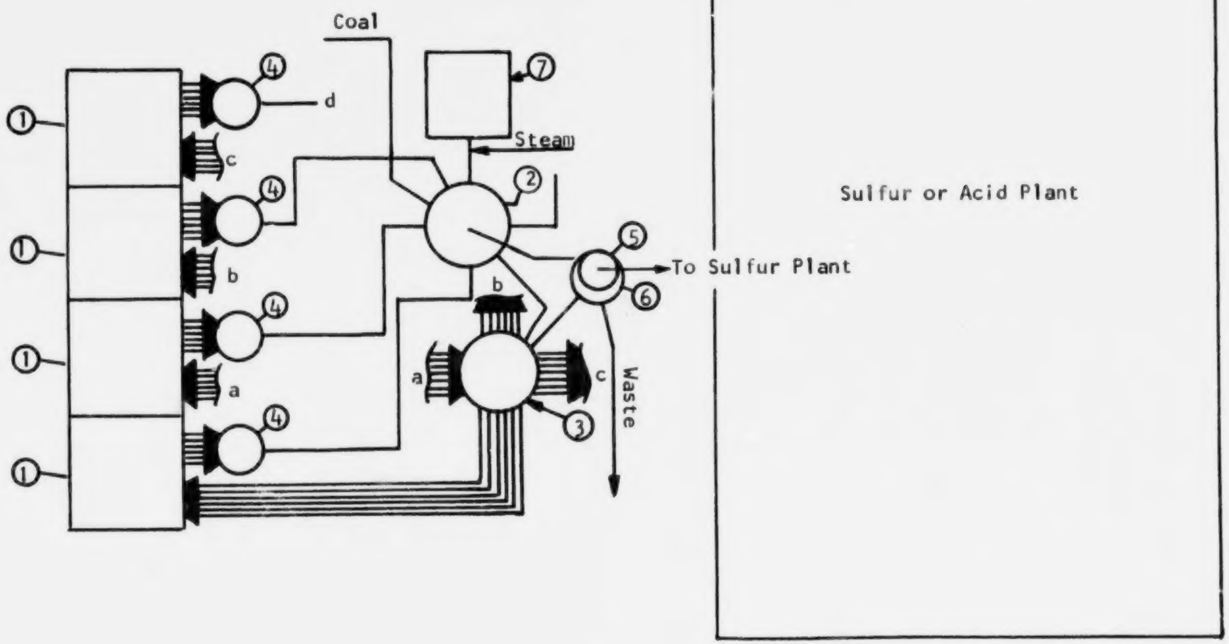


Fig. 1-141 – Atmospheric pressure regenerator plant layout

Regenerated stone flows from the reactor vessel to a distributor vessel. Twenty-four lines return regenerated stone from the distributor to each bed in the boilers. Some deactivated stone is discarded from the distributor vessel to provide for the fresh limestone make-up required to achieve the desired SO_2 removal. Fresh stone is charged to the boilers with coal through the coal feed system.

Experimental results from Esso R&E [37] and Esso England [29] indicate that the SO_2 concentration of gas leaving the regenerator vessel will contain the equilibrium concentration of SO_2, approximately 4% to 10% by volume. The sulfur recovery process selected for this gas is the one recommended by Allied Chemicals [2]. The hot regenerator top gas will pass first through a gas cooler and cleaner. The SO_2 concentration will be enriched to 90% by volume by a dimethylamine (DMA) absorber-stripper process. The concentrated SO_2 stream is reduced to sulfur by the Asarco process (described by Allied Chemicals) using natural gas as the fuel.

The obvious points of sulfur loss in the regenerator are by the carrier gas for the spent and regenerated stone. The expected sulfur recovery efficiency for the DMA process is 98% to 99%. This process consumes soda ash and dilute sulfuric acid. Vented gas from the DMA scrubber should contain 450 ppm SO_2. The Asarco process is expected to recover approximately 92% of the sulfur fed to it. Adding a third Claus reactor to the two planned, or the Claus tail gas clean-up, could possibly improve the performance. The overall sulfur recovery efficiency for the regenerator-sulfur recovery plant should be between 85% and 90%.

Emissions from an acid plant should contain 500 ppm or less SO_2, corresponding to a sulfur recovery efficiency in excess of 99%. The overall plant sulfur recovery should be approximately 95%.

The cost estimate for the regeneration system for a 300 MW plant is $6.3/kw. The cost estimate scaled to 635 MW is $4.2/kw. The sulfur recovery system cost estimate is $11/kw for a 635 MW plant.

Claus plant tail gas clean-up will add an estimated \$3-4/kw. Cost estimates for the various components of this regeneration system were obtained from the following sources.

- Solids feed system: scaled from Petrocarb system data
- Sulfur plant: cost estimates in Allied Chemical reports
- Process vessels: scaled from Westinghouse Heat Transfer Department estimate for high-pressure process vessels using extrapolation procedure given by Guthrie.(1)

Auxiliary Equipment

Coal Handling and Crushing. The coal handling system includes receiving, storage, crushing, and conveying facilities (see Figure 1-142). Coal (1-1/2 in x 0) arrives at the plant by unit train twice a week and is unloaded in about five hours. A 12,000-ton silo acts as a buffer between the batch-type receiving operation and the continuous feeding operation. It also provides live storage for short-term delays in deliveries. Dead storage for long-term delays in deliveries is provided by a 50,000-ton open pile.

Coal from the silo is crushed to 1/4 x 0 in. in a reversible hammermill. The size distribution of this crushed coal is given in Appendix K. The crushed coal is conveyed through a 150-ton surge hopper to a group of 16 bunkers or hoppers which serve the individual feed injectors to the boiler modules.

Make-up Sorbent Handling. The limestone sorbent is purchased crushed to size (1/4 x 0 in) and is delivered in covered cars. A silo is provided with sufficient capacity for both live and dead storage. The crushed sorbent is fed onto the coal conveyor just ahead of the 150-ton surge hopper.

Coal Drying and Feeding. The coal drying and feeding system is shown in Figure 1-143. A direct fired system was chosen over

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(1) H. Popper, Modern Cost Engineering Techniques, New York: McGraw Hill, 1970.

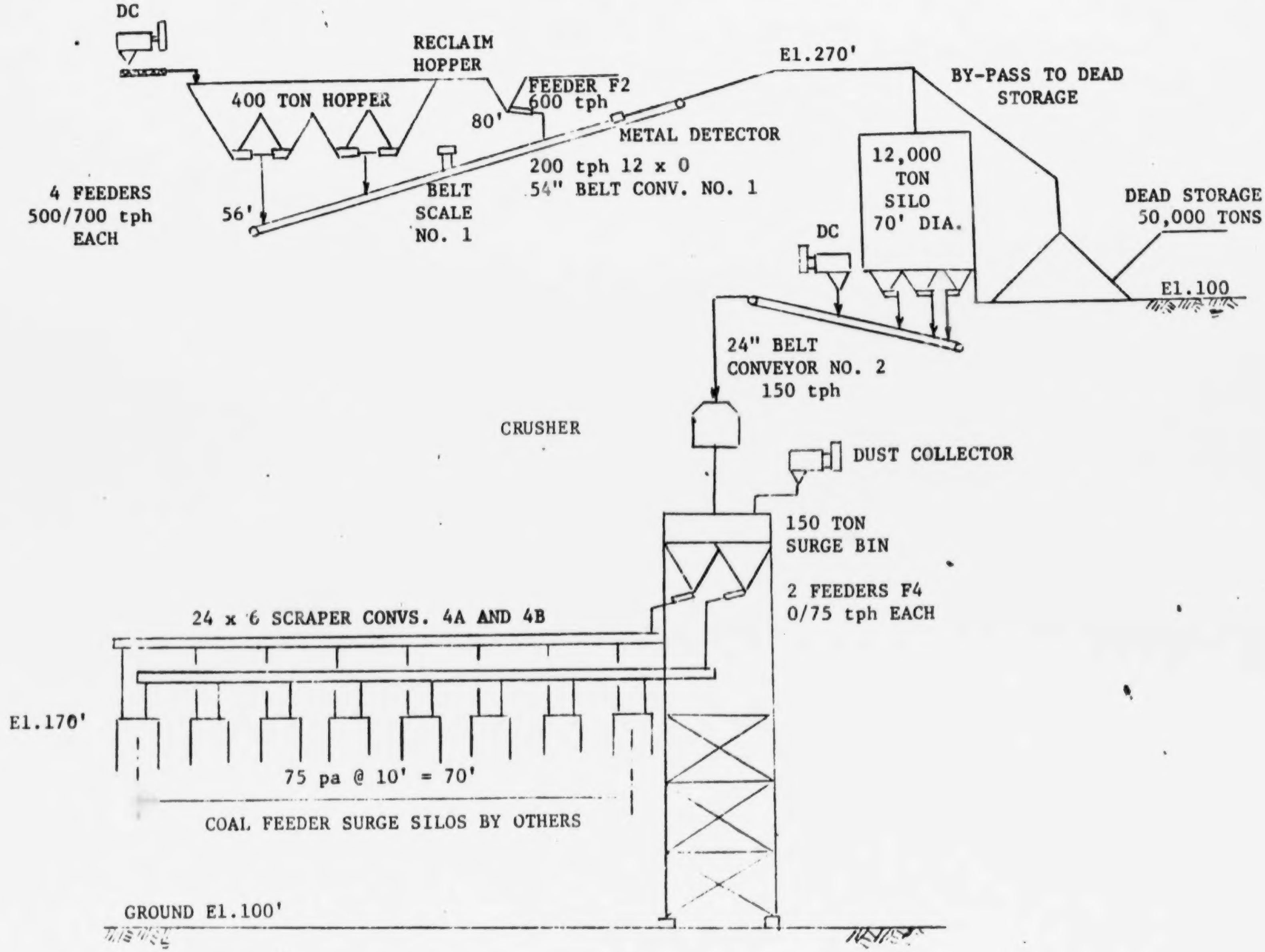

FIG. 1-142 COAL HANDLING PLANT

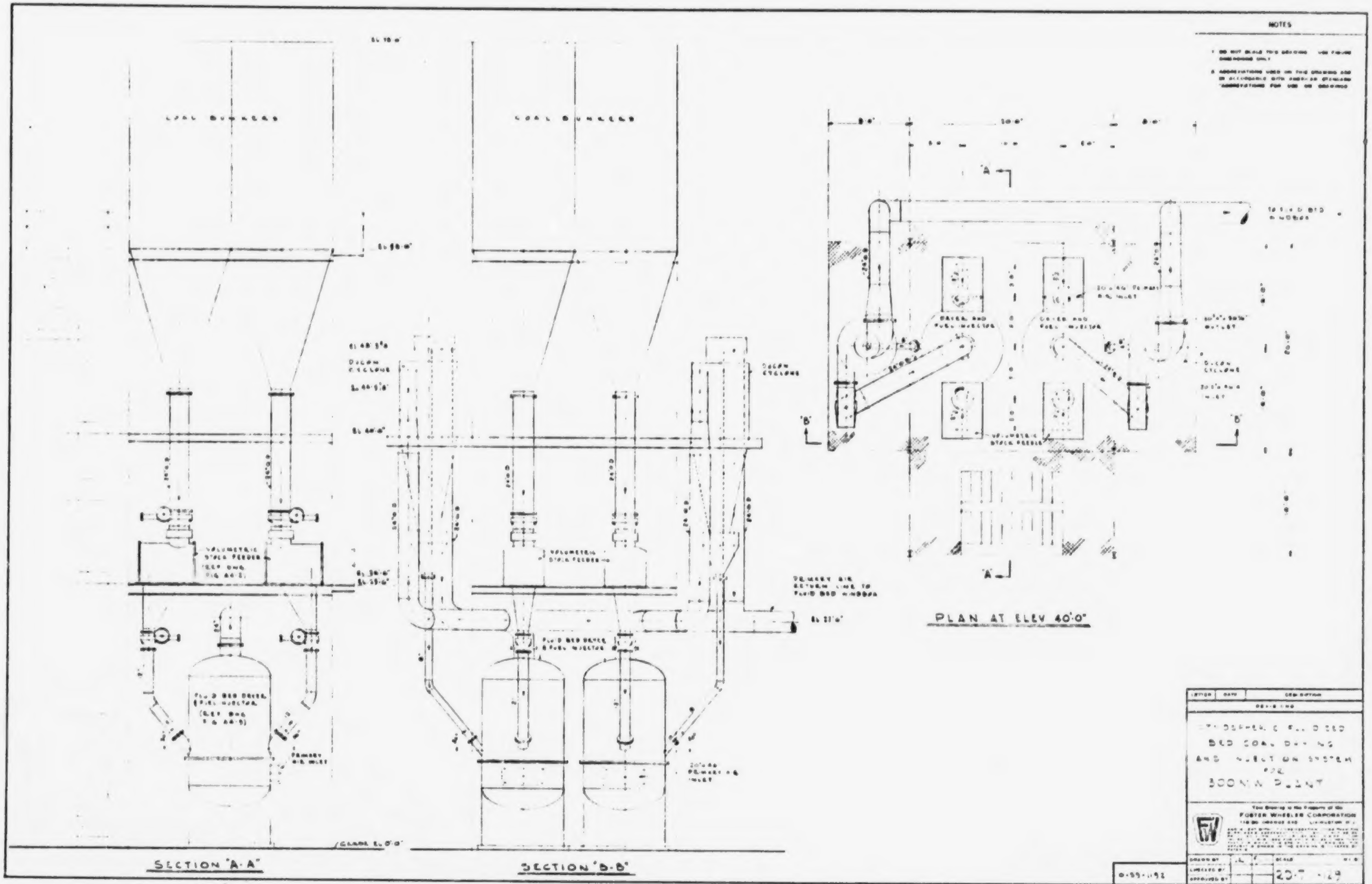

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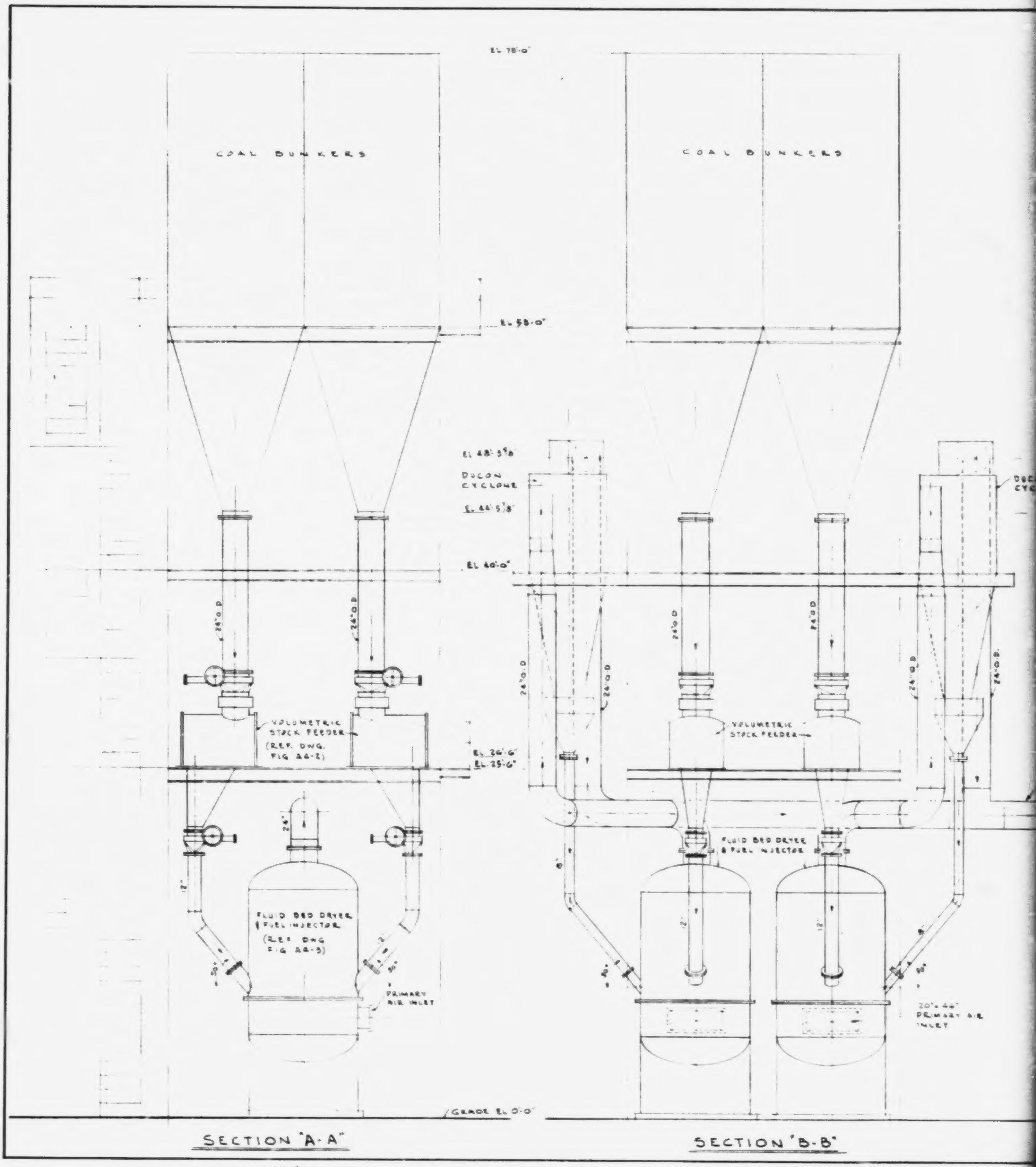
Fig. 1-143

369.2



SECTION "A-A"

SECTION "B-B"

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Fig. 1-143

369.2

NOTES

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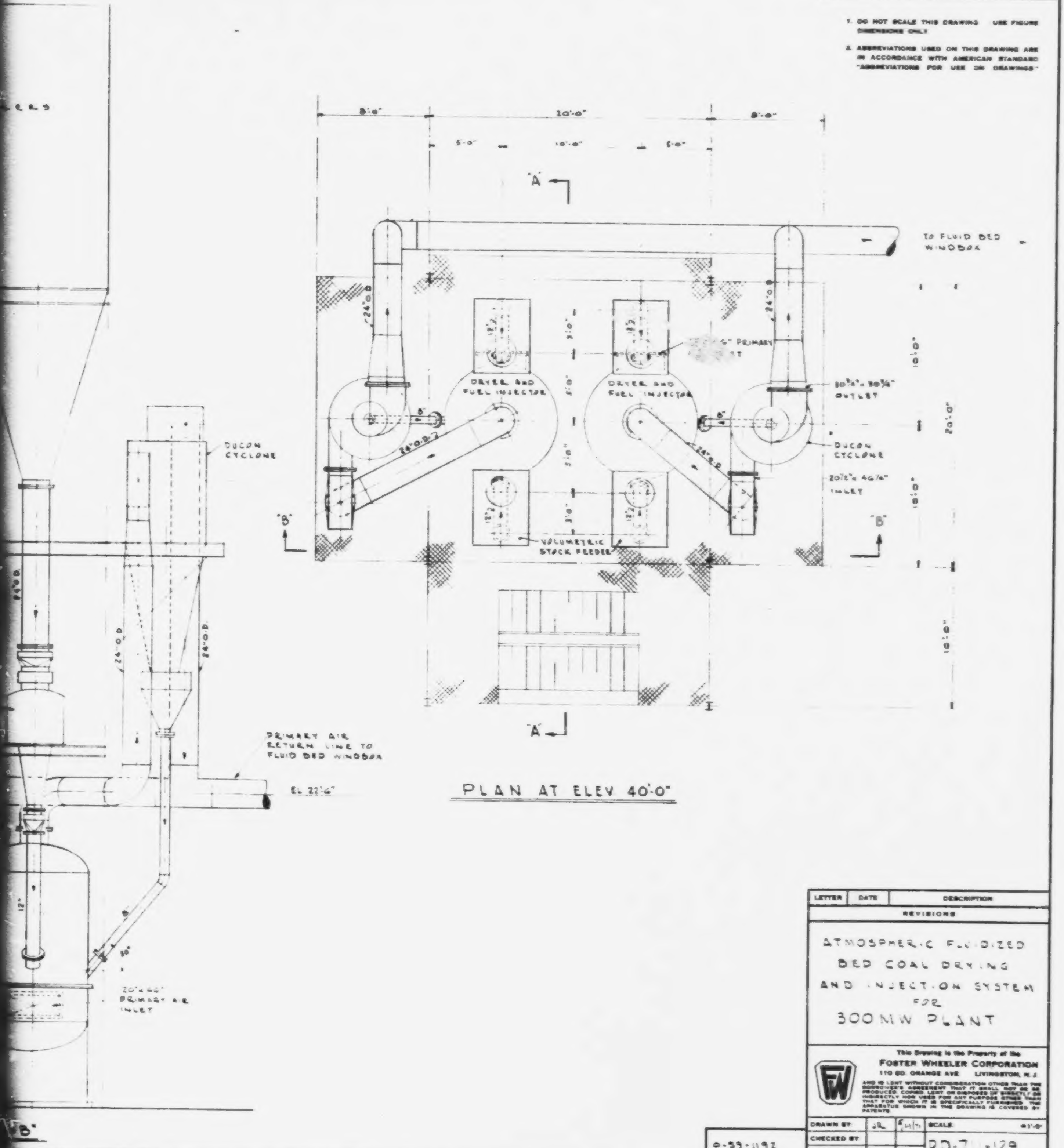

LETTER	DATE	DESCRIPTION
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ATMOSPHERIC FLUIDIZED
BED COAL DRYING
AND INJECTION SYSTEM
FOR
300MW PLANT

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Fig. 1-143

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several alternatives. This feeding system includes bunkers for receiving coal from the coal handling systems, volumetric feeders, combination fluid bed dryers and distributors, dust collectors for the fluidizing air, primary air fans, and dilute phase coal transport lines to the individual coal injection points. The flow diagram for the coal feeding system is shown in Figure 1-144. Heated air from the air preheater goes to the primary air fans for use in the dryer and fuel injector. This air is then cleaned and used as part of the combustion air for the primary beds. Minor amounts of coal transport and sealing air are also required by the coal feeding system.

Carbon Burn-up Cell Feeding. The particulates elutriated from the primary beds are assumed to contain about 13% of the fuel fired. Over 90% of this material is recovered for recycle to the carbon burn-up cell (see Figure 1-145). The recovered material at a temperature of 840°F is collected in a dust hopper from which it is periodically transferred into a fluid bed injector which meters the flow of solids to the feed points in the carbon burn-up cell.

Draft Fans and Air Preheater. A forced draft system was selected over a balanced draft system since the gas-tight fluidized bed boiler construction is well suited to the forced draft system. The secondary air enters the system at the forced draft fan and is heated to 735°F in the air preheater. It is fed from there into the internal air distribution duct. A separate cool stream which is taken off ahead of the air heater goes to the carbon burn-up cell. A tubular type air preheater was selected over the somewhat less expensive regenerative type because of the severe leakage problem which would result from the use of the latter in combination with the high-pressure drop fluidized bed and the forced draft fan.

Particulate Removal. The particulate removal system in the fluidized bed boiler has two functions: to recover most of the unburned fuel in the solids elutriated from the primary beds for recycle to the carbon burn-up cell and to control the emission of particulates to the atmosphere (see Figure 1-146). There are two stages of particulate

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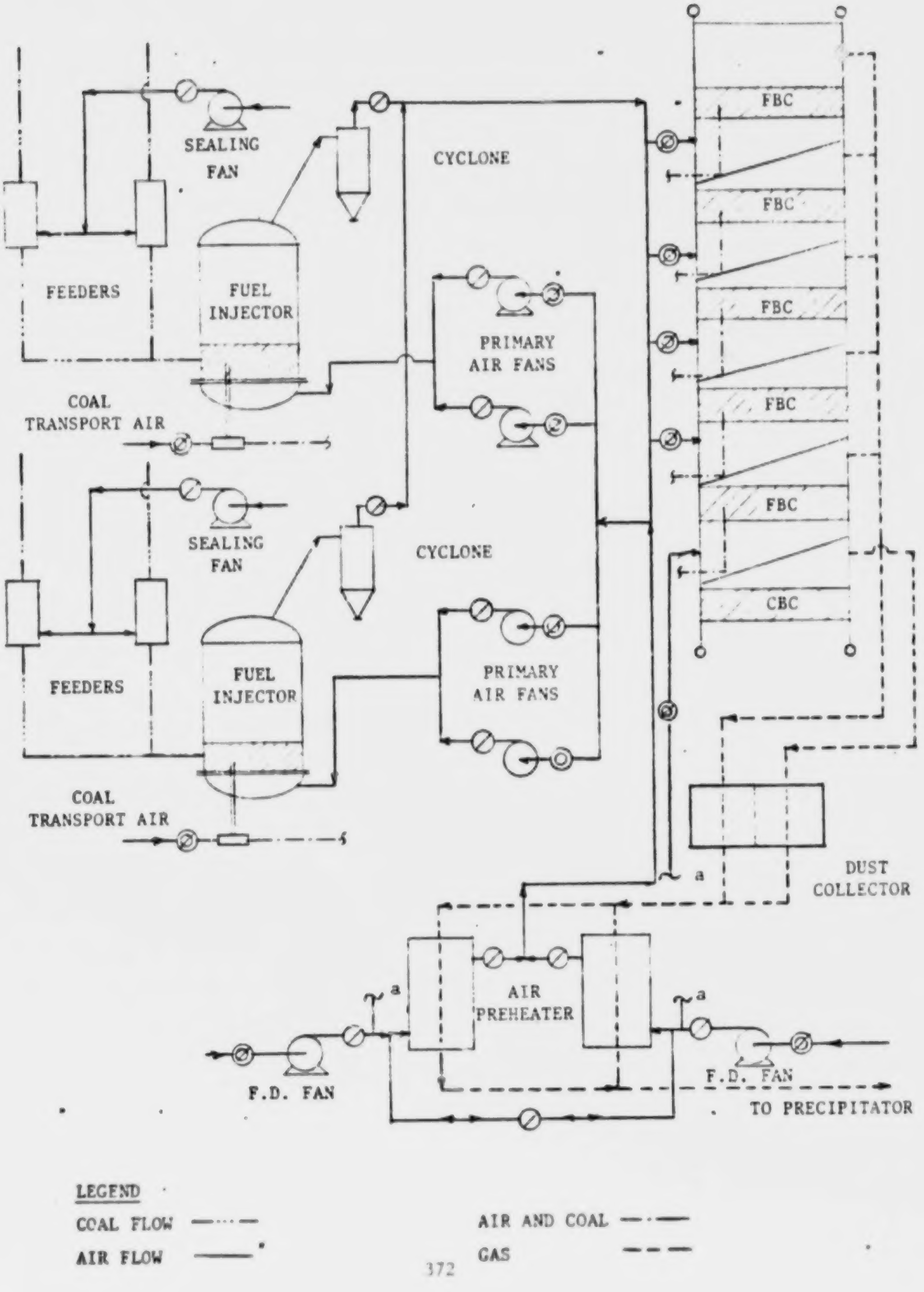
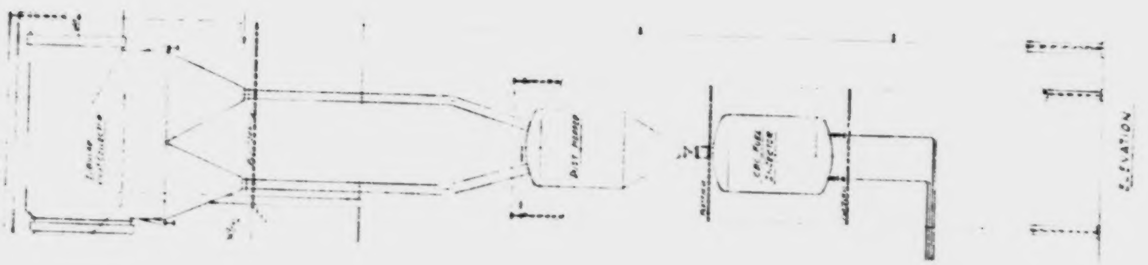
FIG. 1-144 AIR FLOW DIAGRAM





DUST COLLECTOR
ATMOSPHERIC FLUIDIZED
BED STEAM GENERATORS
FOR
300 MW PLANT
FOUR (4) REQUIRED

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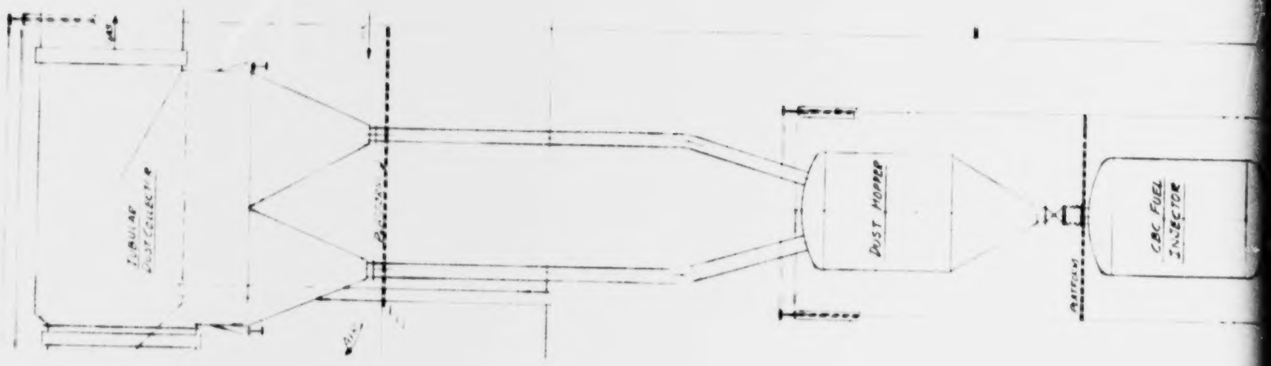
TUBULAR DUST COLLECTOR
DUST HOPPER
CBC FUEL INJECTOR
PLATFORM

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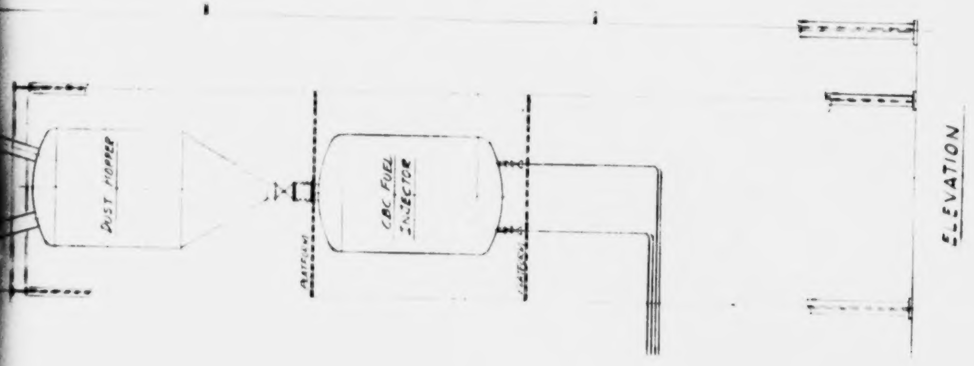

Fig. 1-145

DUST COLLECTOR
ATMOSPHERIC FLUIDIZED
BED STEAM GENERATOR
FOR
300 MW PLANT

FOUR (4) REQUIRED

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removal for the products of combustion from both the primary beds and the carbon burn-up cell.

Multiclone collectors were selected for the first stage of both streams. The two are physically integrated, with the streams isolated so that the carbon-rich ash from the primary beds is not mixed with the spent ash from the carbon burn-up cell. For the predicted particle size distributions from the primary beds and the carbon burn-up cell, the collection efficiencies are estimated to be 96% and 90%, respectively.

The two gas streams are mixed after the multiclones, and an electrostatic precipitator is used to reduce the particulate concentration to less than the specified 0.01 grains per standard cubic foot.

Systems Analysis

The preliminary design for the atmospheric pressure fluidized bed boiler was to have been made for a 300 MW plant. Due to an error in the steam flow specified for this plant, however, the design capacity was 280 MW. A flow sheet for this boiler at full load with solids flows is shown in Figure 1-147, and an energy balance at full load is shown in Figure 1-148. In these mass and energy balances, the combustible solids in the ash elutriated from the primary beds and the carbon burn-up cell were considered to have the same composition as the original coal. In actuality, the unburned solids would be mostly elemental carbon.

A temperature-enthalpy diagram for the atmospheric pressure boiler is shown in Figure 1-149. The theoretical gas temperature, the fluidized bed temperature, and the cold-side fluid temperatures are shown for the full-load conditions. This plot graphically describes the allocation of boiler functions and illustrates the heat transfer situation in each boiler component.

Tables 1-45, 1-46, and 1-47 give the fluidized bed parameters for one of four boiler modules for 100%, 75%, and 50% boiler loads.

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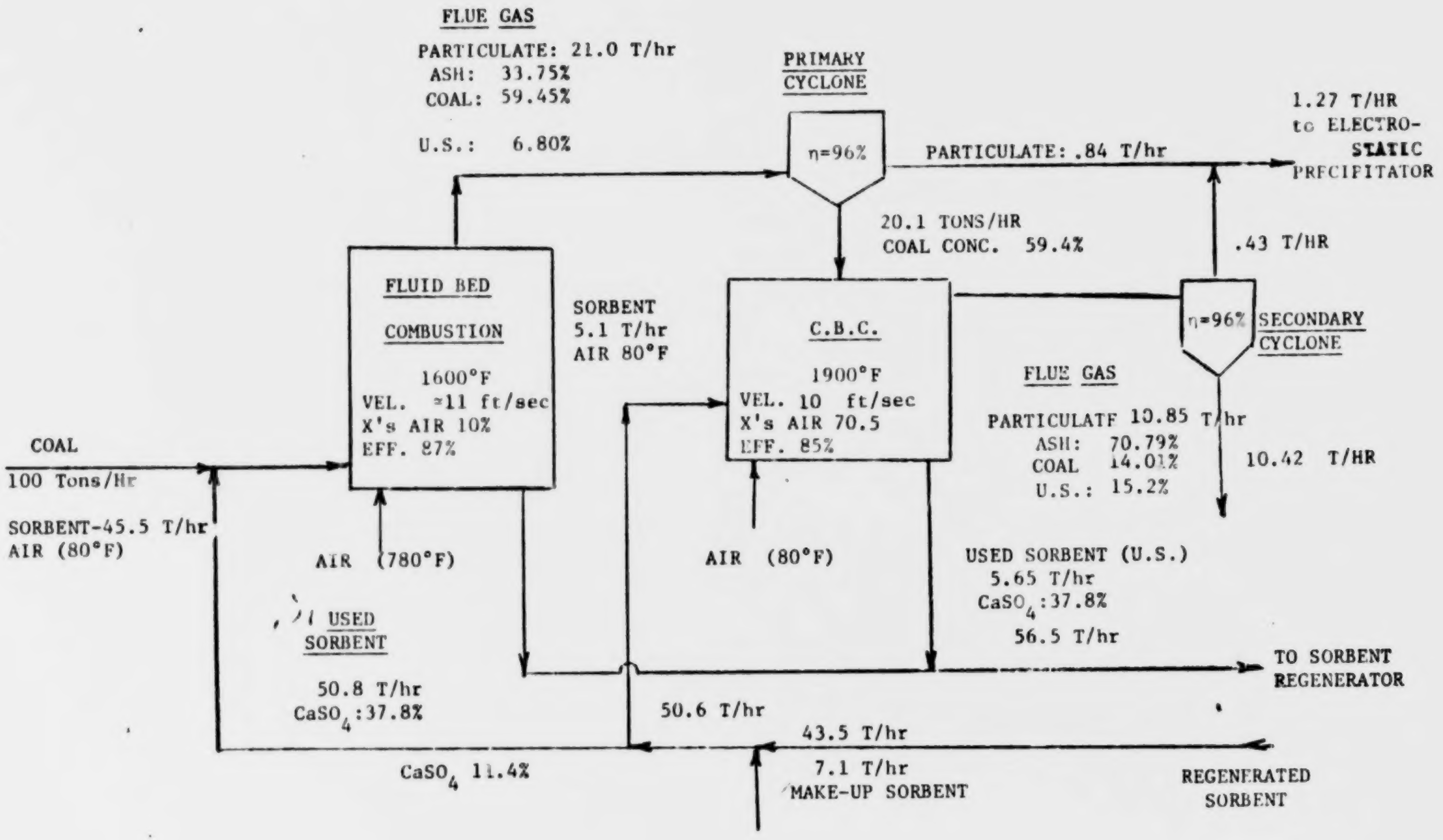

FIG. 1-147 OVERALL MATERIAL BALANCE--FULL LOAD

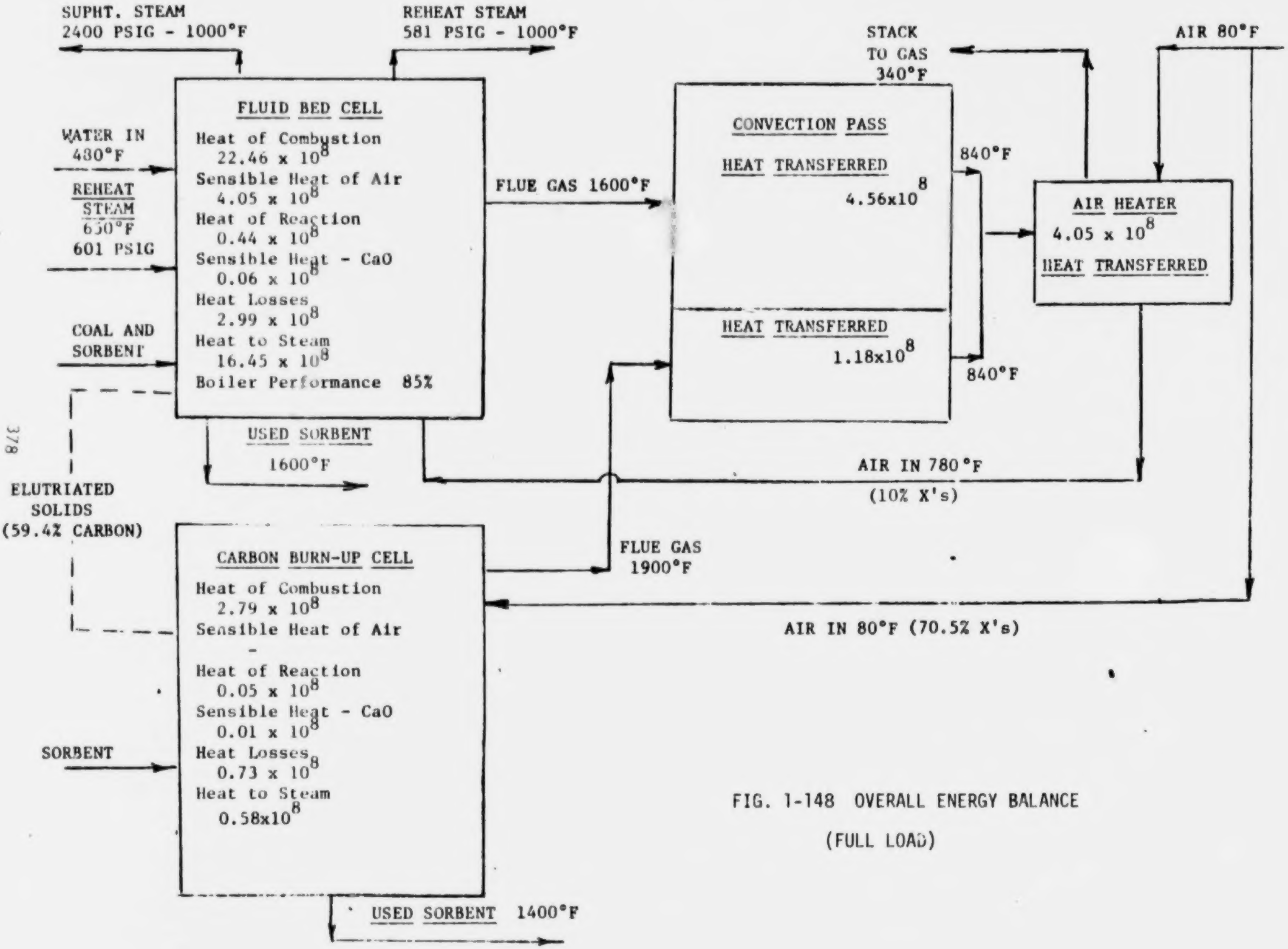

FIG. 1-148 OVERALL ENERGY BALANCE

(FULL LOAD)

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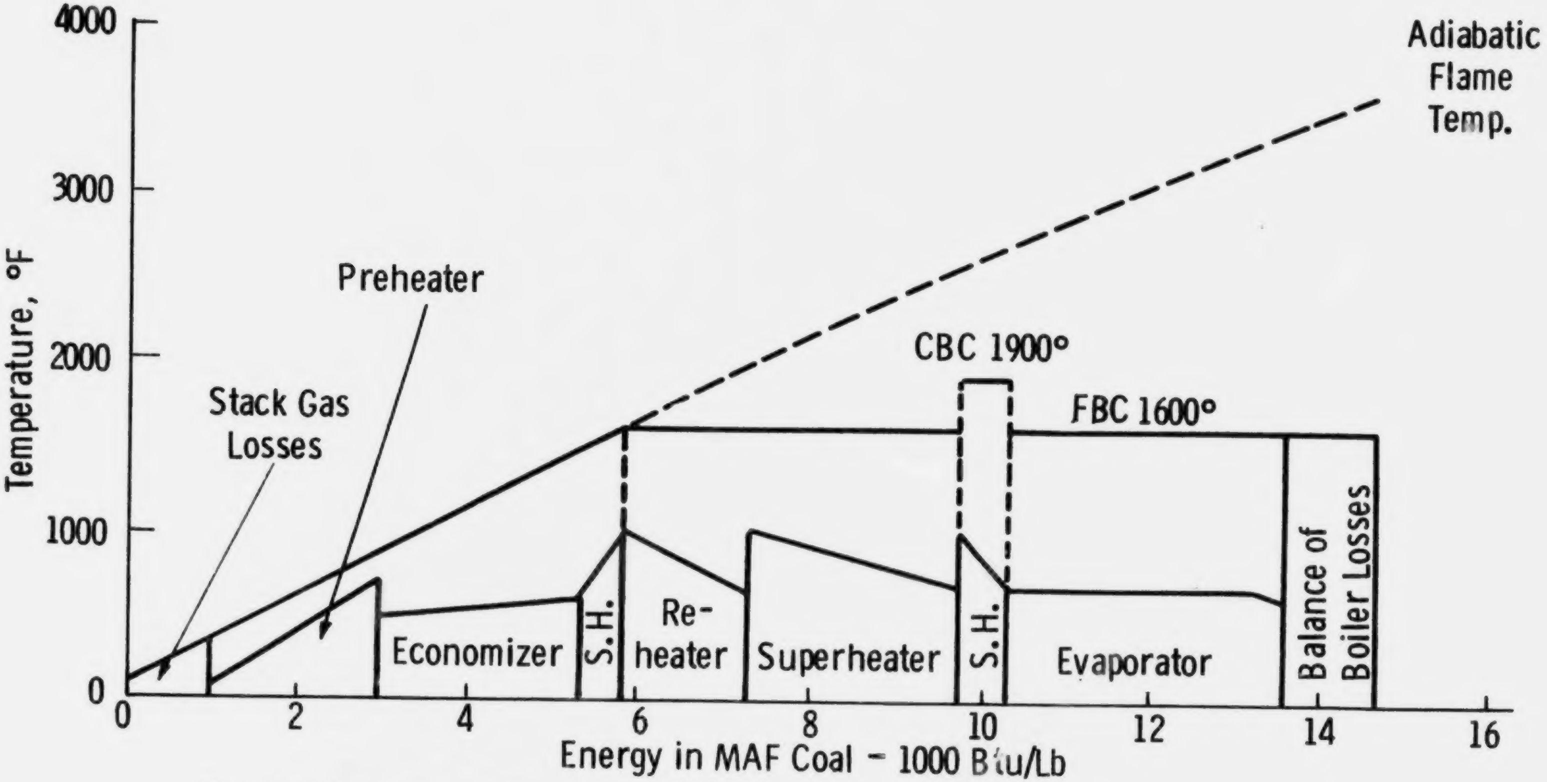

Fig. 1-149—Temperature-energy diagram for the atmospheric utility boiler design

TABLE 1-45

FLUIDIZED BED PARAMETERS

DESIGN LOAD

(ONE MODULE)

	FUEL FLOW LB/HR	AIRFLOW LB/HR	FLUE GAS LB/HR	SUPERFICIAL VEL. FT/SEC	BED TEMPERATURE °F
REHEATER	11,600(a)	$0.126\text{x}10^6$	$0.137\text{x}10^6$	10.9	1600(d)
SUPERHEATER I	11,700(a)	0.126	0.138	11.0	1600(d)
SUPERHEATER II	11,700(a)	0.126	0.138	11.0	1600(d)
EVAPORATOR I	7,000(a)	0.076	0.083	7.85	1600(d)
EVAPORATOR II	7,000(a)	0.076	0.083	7.85	1600(d)
CBC	10,300(b)	0.103(c)	0.108	10.7	1900(d)

(a) Fuel as fired

(b) 59.45% coal
33.75% ash
6.80% sorbent

(c) 70.5% X's air based on coal flow to CBC

(d) Specified by Westinghouse

TABLE 1-46

FLUIDIZED BED PARAMETERS

75% BOILER LOAD

(ONE MODULE)

	FUEL FLOW LB/HR	AIRFLOW LB/HR	FLUE GAS LB/HR	SUPERFICIAL VEL. FT/SEC	BED TEMPERATURE °F
REHEATER	8,630(a)	$0.0935\text{x}10^6$	$0.102\text{x}10^6$	8.10	1425
SUPERHEATER I	8,790(a)	0.095	0.104	7.91	1390
SUPERHEATER II	8,790(a)	0.095	0.104	8.29	1433
EVAPORATOR	5,250(a)	0.057	0.062	4.74	1350
EVAPORATOR II	5,250	0.057	0.062	4.74	1350
CBC	7,670(b)	0.076(c)	0.081	6.10	1900(d)

(a) Fuel as fired

(b) 59.45% coal
33.75% ash
6.80% sorbent

(c) 47% X's air based on coal flow to CBC

(d) Specified by Westinghouse

TABLE 1-47

FLUIDIZED BED PARAMETERS

50% BOILER LOAD

(ONE MODULE)

	FUEL FLOW LB/HR	AIRFLOW LB/HR	FLUE GAS LB/HR	SUPERFICIAL VEL. FT/SEC	BED TEMPERATURE °F
REHEATER	5,800[(a)]	0.063×10^6	0.068×10^6	4.80	1223
SUPERHEATER I	5,050[(a)]	0.064	0.069	4.79	1170
SUPERHEATER II	5,850[(a)]	0.064	0.069	4.70	1218
EVAPORATOR I	3,500[(a)]	0.038	0.041	2.79	1144
EVAPORATOR II	3,500[(a)]	0.038	0.041	2.79	1144
CBC	5,150[(b)]	0.052[(c)]	0.054	5.40	1900[(d)]

(a) Fuel as fired

(b) 59.45% coal
33.75% ash
6.80% sorbent

(c) 21.4% X's air based on coal flow to CBC

(d) Specified by Westinghouse

This shows the variation in the temperatures of the primary beds required at part load to match the functional duty requirements.

Boiler Performance. The losses for the atmospheric pressure fluidized bed boiler are as follows:

Losses	Percent
Dry gas loss (stack temp - 275°F)	4.28
Loss due to hydrogen and moisture in coal	5.05
Loss due to moisture in air	0.10
Radiation loss	0.18
Incomplete combustion	2.39
Unaccounted for and manufacturer's margin	1.50
	13.88

Radiation, moisture, and unaccounted for losses and manufacturer's margin are the same as for a conventional boiler. Dry gas loss is slightly lower than that of most units because of the lower stack gas temperature. The losses due to incomplete combustion are higher because of the elutriation of unburned carbon from the fluidized beds. This gives a boiler efficiency of 86.1%, which is significantly lower than that for the conventional coal-fired boilers.

Plant Performance. The boiler efficiency at design capacity has been estimated to be 86.1%. Figure 1-150 shows the part-load boiler efficiency, which was patterned after that of a conventional boiler. Using this boiler efficiency gives the plant heat rate relationship shown in Figure 1-151. This is two to three percent poorer than that for a conventional plant because of the lower boiler efficiency.

Load Control. A maximum primary bed temperature of 1600°F was chosen on the basis of ash characteristics and sulfur recovery considerations; a lower limit of 1300°F for continuous operation was chosen on the basis of sulfur recovery and combustion efficiency considerations. At part load, the gas-side heat transfer coefficient is nearly constant, and there is only a small reduction in steam-side

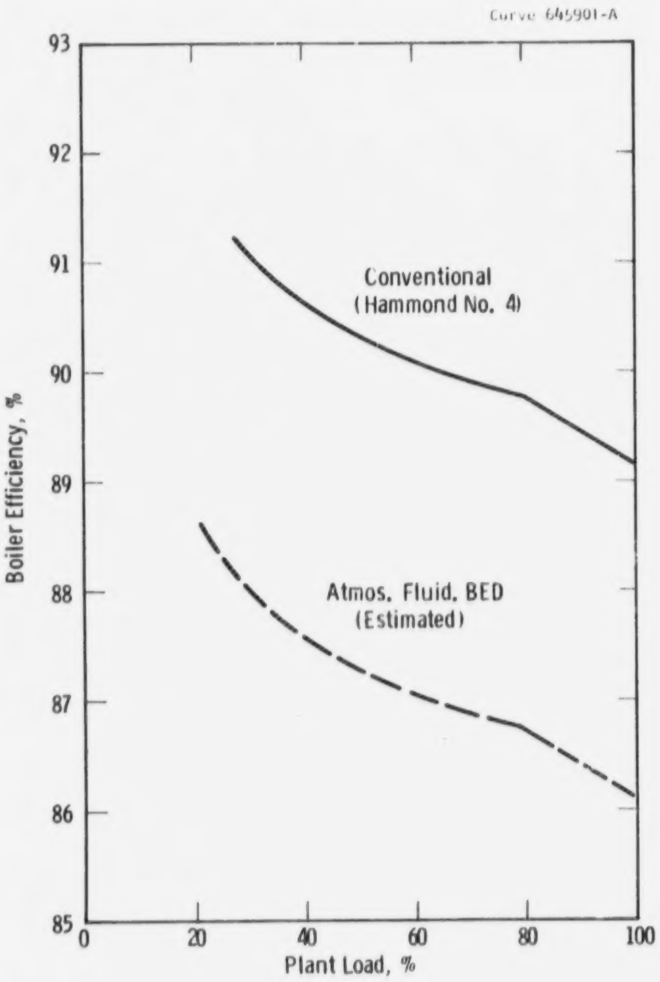

Fig. 1-150 Part load boiler efficiency for conventional and fluid bed boiler

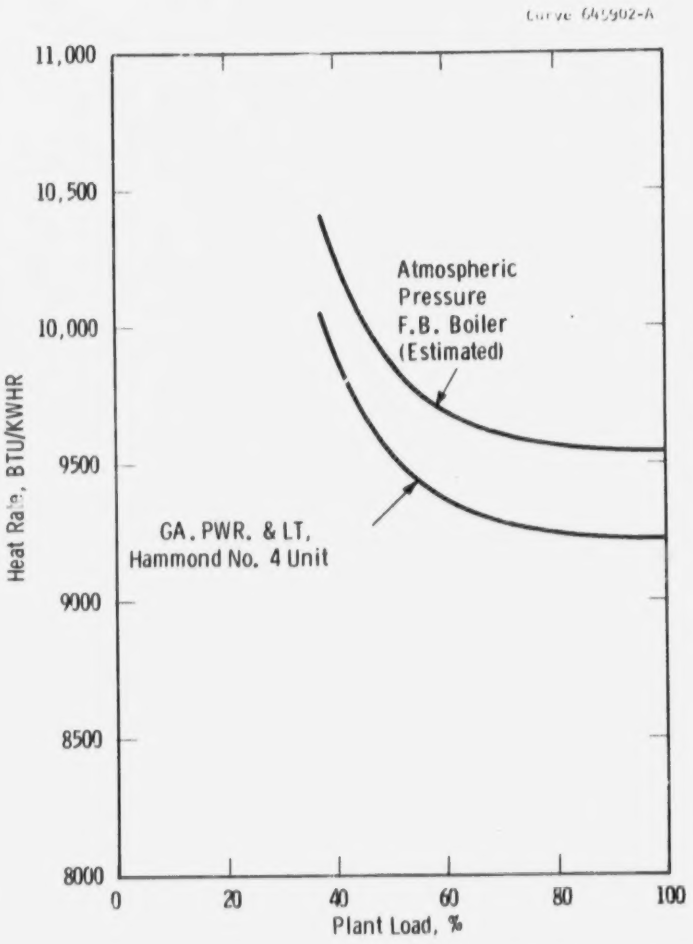

Fig. 1 151 Heat rate vs plant load for conventional and F.B. Boiler 2400/1000/1000/3 in.

coefficient because of Reynolds number effect. The heat transferred from the beds, therefore, is primarily a function of the bed temperature. Because of the constraints on bed temperatures discussed above, the turn-down capability of a boiler module is only 35% of boiler flow, which is equivalent to about 40% reduction in plant power. To attain a 4:1 turn-down capability for the plant it is necessary to use four boiler modules which can be shut down individually. This gives the plant load profile shown in Figure 1-152.

Ignition. Ignition of the primary beds is done in three steps:

1. Warm-up of bed and pressure parts
2. Heating of bed to ignition temperature
3. Injection of coal at ignition temperature.

The ignition temperature of the specified coal is about 850°F. Six 6×10^6 BTU/hr burners located in the air ducts between the air preheater and the boiler module are used to warm up the beds and the pressure parts. During step 1 the airflow is increased until the bed is fluidized and the bed temperature approaches 650°F. Then a set of igniter burners located immediately above the bed surface is turned on to heat the bed to the ignition temperature. The igniter burners are rated at 2×10^6 BTU/hr.

The carbon burn-up cell will receive less reactive fuel, so it will require a higher ignition temperature. Since ignition of the carbon burn-up cell will lag behind that of the primary beds, high-temperature air will be available for preheating the bed. The carbon burn-up cells will be equipped with an above-the-bed igniter burner with off-on control sized to provide the heat necessary for ignition.

Start-up. The start-up of the atmospheric pressure fluidized bed boiler is the same as that of a conventional once-through boiler (see Appendix K), but with the modular boiler only one module is started up at a time. The start-up of the second, third, and fourth modules will differ slightly from that of the first module since feedwater

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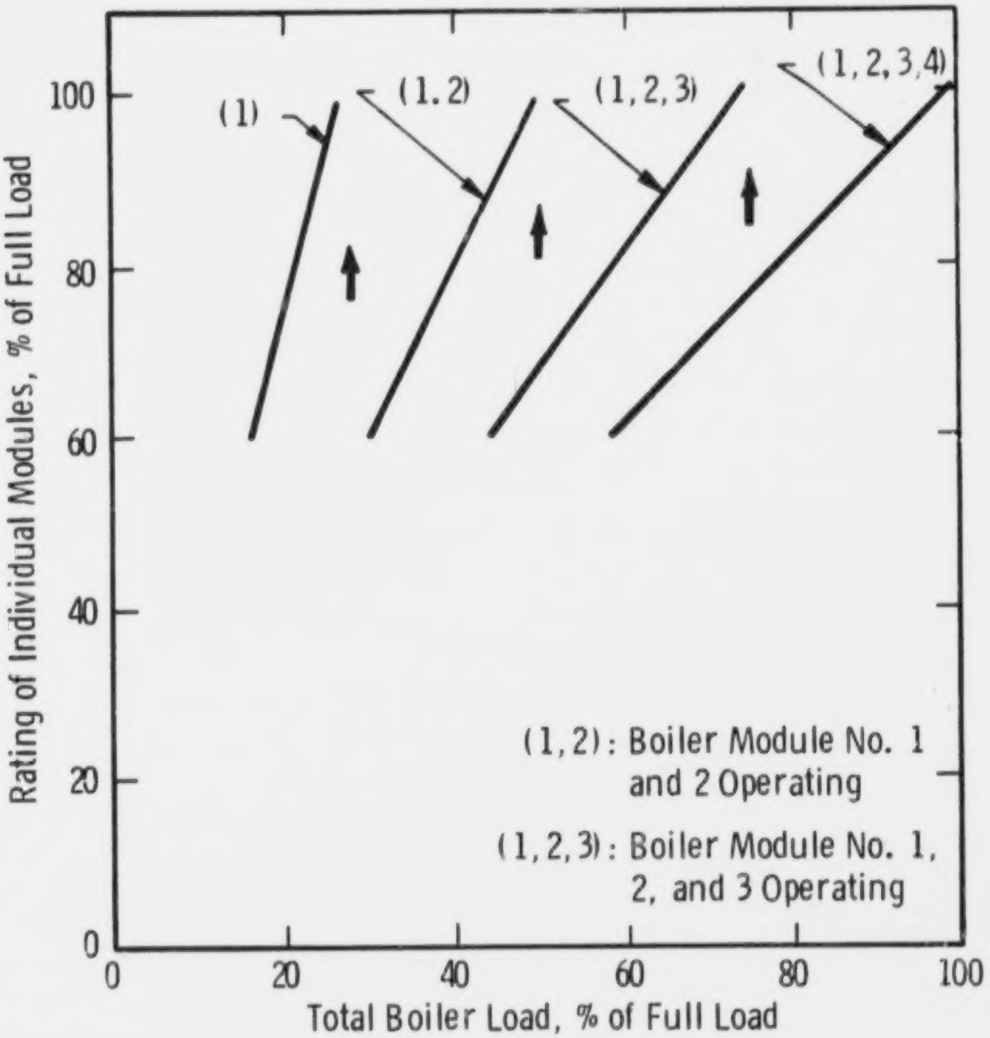

Fig. 1-152 — Boiler load reduction

will be at 500°F rather than 80°F. Details of this start-up procedure are given in Appendix K.

Shut-down. Normal shut-down is essentially a reversal of the start-up procedure. Fuel flow to individual modules would be cut off, but fluidizing air would be maintained to purge and cool the beds. The air would be shut off when the bed was sufficiently cooled. If it was desirable to keep the bed in stand-by condition, feedwater flow at about 500°F would be continued through the pre-evaporator, the evaporator walls, and the first superheater.

Emergency shut-down due to steam-side problems is accomplished in the same manner as in conventional units. The fluidized beds do have a high thermal inertia and long residence time for the fuel. The bed temperatures are low, however, and it is estimated that the temperature excursion resulting from interruption of steam flow will not exceed tube material limitations.

Pressure Losses. The gas side pressure losses at full load are as follows:

COMPONENT		PRESSURE LOSS IN H_2O
Ducts		1.43
Distribution plate		10.8
Bed		27.5
Convection bank		0.1
Dust collectors		3.0
Air heaters		
Air-side		3.0
Gas-side		1.5
	TOTAL	47.33

Since the bed pressure loss is predominant and independent of load, the overall gas-side pressure drop will be relatively unchanged with load.

The steam-side pressure losses are as follows:

COMPONENT		PRESSURE LOSS-PSI
Primary Loop		
Feed heating, convection bank		125
Evaporator bundle 1		96
Evaporator bundle 2		
Water walls pass 1		41
Water walls pass 2		55
Water walls pass 3		55
Water walls pass 4		51
Superheater bank 1		142
Superheater bank 2		
	SUBTOTAL	565
Reheater Loop		17
	TOTAL	582

Economics

Capital and energy costs were prepared for pressurized and atmospheric pressure fluidized bed boiler power plants burning high-sulfur coal based on the preliminary designs. Capital and energy costs were also estimated for fluidized bed boiler power plants burning high-sulfur oil. Costs are projected for plants with once-through limestone/dolomite and with stone regeneration and sulfur recovery. These costs are compared with the cost of conventional coal- and oil-fired plants with stack gas cleaning.

Assumptions common to all plants are:

- Northeast site
- January 1970 costs
- Begin construction July 1971
- Supplementary cooling not included
- Contingency of 6%
- Escalation at 7-1/2%
- Interest during construction at 7-1/2%
- 70% capacity factor
- 15% levelized fixed charge
- No credit taken for sulfur recovery
- 45¢/10^6 Btu high-sulfur coal (∿ 4%)
- 45¢/10^6 Btu high-sulfur oil (∿ 3%)
- $2/ton limestone or dolomite.

The heat rates for each coal-fired plant are:

Conventional	9230 Btu/kwh
Atmospheric pressure fluidized bed boiler	9550 Btu/kwh
Pressurized fluidized bed boiler combined cycle	8967 Btu/kwh

The heat rates are for 3 in. Hg turbine back pressure to allow for a cooling tower.

Capital and energy costs for the plants are for 635 MW capacity. The costs assume each concept has been developed and thus represent commercial costs. Costs as a function of plant capacity were projected using conventional plant characteristic cost curves.

Capital Costs

Capital costs for 635 MW coal-fired plants are summarized in Table 1-48. Boiler plant equipment and steam turbine plant equipment component costs are presented in Tables 1-49 and 1-50. Capital cost as a function of plant capacity is shown in Figure 1-153. Air pollution control costs are summarized in Table 1-51.

Pressurized Combined Cycle Plant. The total capital cost estimate for the plant with sulfur recovery is $276/kw. This represents an 18% cost reduction over a conventional plant with stack gas cleaning.

The cost estimate assumes no advances in subsystem concepts (e.g., solids handling, particulate removal) beyond current technology. Cost reductions may be achieved by using alternative subsystem concepts. The following components have been studied for potential savings:

- Particulate Removal: The projected system utilizes four tornado collectors for final particle removal before the gas turbine. Two alternative systems have been considered which reduce the number and size of the units. Cost estimates for these systems show cost reductions of up to $5/kw(1) for the particulate removal equipment.
- High-Temperature Gas Piping: The high-temperature gas piping cost would be reduced if the particulate removal system were simplified by using two tornado units per module for the 600 MW plant. Additional savings might be realized if high-alloy steel is not required between the

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(1) Based on quotation from Aerodyne.

TABLE 1-48

BOILER PLANT EQUIPMENT COSTS

	CONVENTIONAL P.F. BOILER $/KW	ATMOSPHERIC-PRESSURE FLUID BED BOILER $/KW		PRESSURIZED FLUID BED BOILER $/KW	
	Once-Through	Regeneration	Once-Through	Regeneration	Once-Through
Steam Generator	31.50	25.10	25.10	14.49	14.49
Draft System					
F.D. Fans	0.98	1.50	1.50	--	--
Particulate Removal	(a)	5.00	5.00	12.76	12.76
Draft Flues & Ducts	2.63	1.66(b)	1.66	1.66	1.66
Piping (to gas turbines)	--	--	--	2.77	2.77
Stack and Foundation	0.47(c)	0.47	0.47	0.47	0.47
Air Heater	3.30(d)	4.63	4.63	--	--
Coal Handling & Feeding Equipment	13.58	11.50(e)	14.71	10.94(e)	14.94
Ash & Dust Handling Systems	2.69	2.69(f)	3.69	0.55(f)	1.55
Stack Gas Cleaning System	20.0 (g)	--	--	--	--
Regeneration System	--	4.2	--	8.90	--
Sulfur Recovery System	--	11.0 (h)	--	3.78(h)	--
[illegible] Control	3.50	--	--	--	--
Instrumentation & Controls(i)	2.64	3.10	3.10	3.10	3.10
Miscellaneous Equipment	0.94	0.94	0.94	0.94	0.94
	82.23	71.79	60.80	60.36	52.68

(a) No particulate control assumed beyond wet scrubber. (see note 9)

(b) Ducts incorporated with steam generator; cost between air preheaters, precipitator, and stack obtained from United Engineers cost estimate for ducting at the same conditions for the pressurized plant.

(c) Stack height assumed for all plants of 280 ft based on emissions projected from pressurized fluid bed boiler combined cycle plant.

(d) Air heater cost supplied by Foster Wheeler.

(e) The coal handling system costs for both systems were prepared by McNally-Pittsburg. The coal feeding system cost for the pressurized system was prepared by Petrocarb and United Engineers and Constructors. The coal feeding system equipment cost for the atmospheric-pressure system was prepared by Foster-Wheeler. The cost for foundation and erection of the system was added. Differences in the coal feed systems include: atmospheric system has four times the coal feed pipes and includes coal drying as part of the injector system which requires a cyclone system for each injector; pressurized system has one lock hopper per injector, atmospheric system has two volumetric feeders per injector; atmospheric system includes primary air fans, pressurized system includes booster compressor. (Pressurized system coal drying included with coal handling cost.)

(f) The atmospheric ash handling system is assumed to be the same as a conventional plant. The cost for the pressurized ash handling system was prepared by United Engineers and Constructors. The pressurized system is able to utilize a dry conveying and storage system and thus only includes a conveying system from the boiler modules and regeneration systems to the ash silo. The ash silo and compressor costs are included elsewhere. The atmospheric-pressure systems use wet transport which is more expensive due to the extra equipment -- ash sluice pumps, ash hoppers, dewatering bins, ash settling basin, exhausters, silo, etc.

(g) Recent cost data presented at the International Symposium for West Lime/Limestone Scrubbing, Nov. 1971, indicate the cost for SO_2 and particulate control equipment for a new plant will be ~ $40/kW. Approximately half for SO_2 control and half for particulate control.

(h) Recent information from Ford, Bacon, and Davis indicate they have developed modifications to the sulfur recovery process which reduces sulfur emissions from the process. These modifications will reduce sulfur emissions and increase their reported cost estimate.

(i) Instrumentation for stack gas cleaning, regeneration, and sulfur recovery systems is included with the respective systems.

TABLE 1-49

COAL-FIRED POWER PLANT COST BREAKDOWN
635 MW Plant
(Sulfur Removal Process(es) Designated Under Concept)

	CONVENTIONAL P.F. BOILER(a) $/KW	ATMOSPHERIC-PRESSURE FLUID BED BOILER(b) $/KW		PRESSURIZED FLUID BED BOILER COMBINED CYCLE(c) $/KW	
	Once-Through	Regeneration	Once-Through	Regeneration	Once-Through
Land & Land Rights	1.13	1.13	1.13	1.13	1.13
Structures & Improvements	27.28	25.80	24.80	19.57	18.52
Boiler Plant Equipment	82.23	71.79	60.80	60.36	52.68
Gas Turbine-Generator Equipment	--	--	--	14.80	14.80
Steam Turbine-Generator Plant Equipment	55.76	54.96	54.96	44.14	44.14
Electric Plant Equipment	14.99	14.99	14.99	15.91	15.91
Misc. Plant Equipment	4.21	4.21	4.21	3.56	3.56
Undistributed Costs(d)	30.01	29.20	29.20	28.35	28.35
Other Plant Costs(e)	2.91	2.91	2.91	2.91	2.91
Sub-Total	218.52	204.99	193.00	190.73	182.00
Normal Contingency	13.11	12.30	11.58	11.44	10.92
Sub-Total	213.63	217.29	204.58	202.17	192.92
Escalation	52.12	48.08	45.26	37.60	36.17
Sub-Total	283.75	265.37	249.84	239.77	229.09
Interest During Construction	47.88	41.80	39.35	31.30	30.07
General Items & Engineering	5.5	5.5	5.5	5.5	5.5
TOTAL CAPITAL COST	337.13	312.67	294.69	276.57	264.66

Construction time:

conventional plant	4.5 years(f)
atmospheric pressure fluid bed boiler plant	4.2 years(g)
pressurized fluid bed boiler plant	3.5 years(h)

(a) Conventional plant cost data provided by United Engineers and Constructors. Plant includes throw-away wet scrubbing system and NO_x control.

(b) Cost estimate prepared by Westinghouse on component data supplied by Westinghouse, Foster Wheeler, and United Engineers and Constructors. Plant includes sorbent regeneration and sulfur recovery system.

(c) Cost estimate prepared by United Engineers and Constructors based on component data supplied by Westinghouse, Foster Wheeler, and United Engineers and Constructors. Plant includes sorbent regeneration and sulfur recovery.

(d) Engineering and construction management and supervision, temporary facilities, construction services, etc.

(e) Operator training, spare parts, etc.

(f) Source -- United Engineers and Constructors.

(g) Construction time reduction for boiler projected by Foster Wheeler assumed for total plant.

(h) Construction schedule prepared by United Engineers and Constructors.

TABLE 1-50

STEAM TURBINE PLANT EQUIPMENT COSTS

	CONVENTIONAL P.F. BOILER $/KW	ATMOSPHERIC PRESSURE FLUID BED BOILER $/KW	PRESSURIZED FLUID BED BOILER COMBINED CYCLE(a) $/KW
Turbine generator	22.60	22.60	18.69
Circulating water system	5.46	5.46	4.11
Condensing system	7.33	7.33	5.46
Feedwater system(b)	13.35	12.96(c)	10.01(c)
Other turbine plant equipment(b)	5.57	5.16(c)	4.63(c)
Instruments and controls	1.45	1.45	1.24
TOTAL	55.76	54.96	44.14

(a) Power from steam turbine-generator plant is ∿ 82.6% of total power.

(b) Includes station piping -- boiler feed piping, main steam piping. Includes stack gas coolers in pressurized system.

(c) United Engineers and Constructors developed preliminary piping schematics for the pressurized system and estimated the cost. The small, modular pressurized boilers reduced the piping distance and allowed for smaller, more flexible, pipe which reduced the piping cost. The atmospheric-pressure boiler piping cost was estimated from the pressurized system cost.

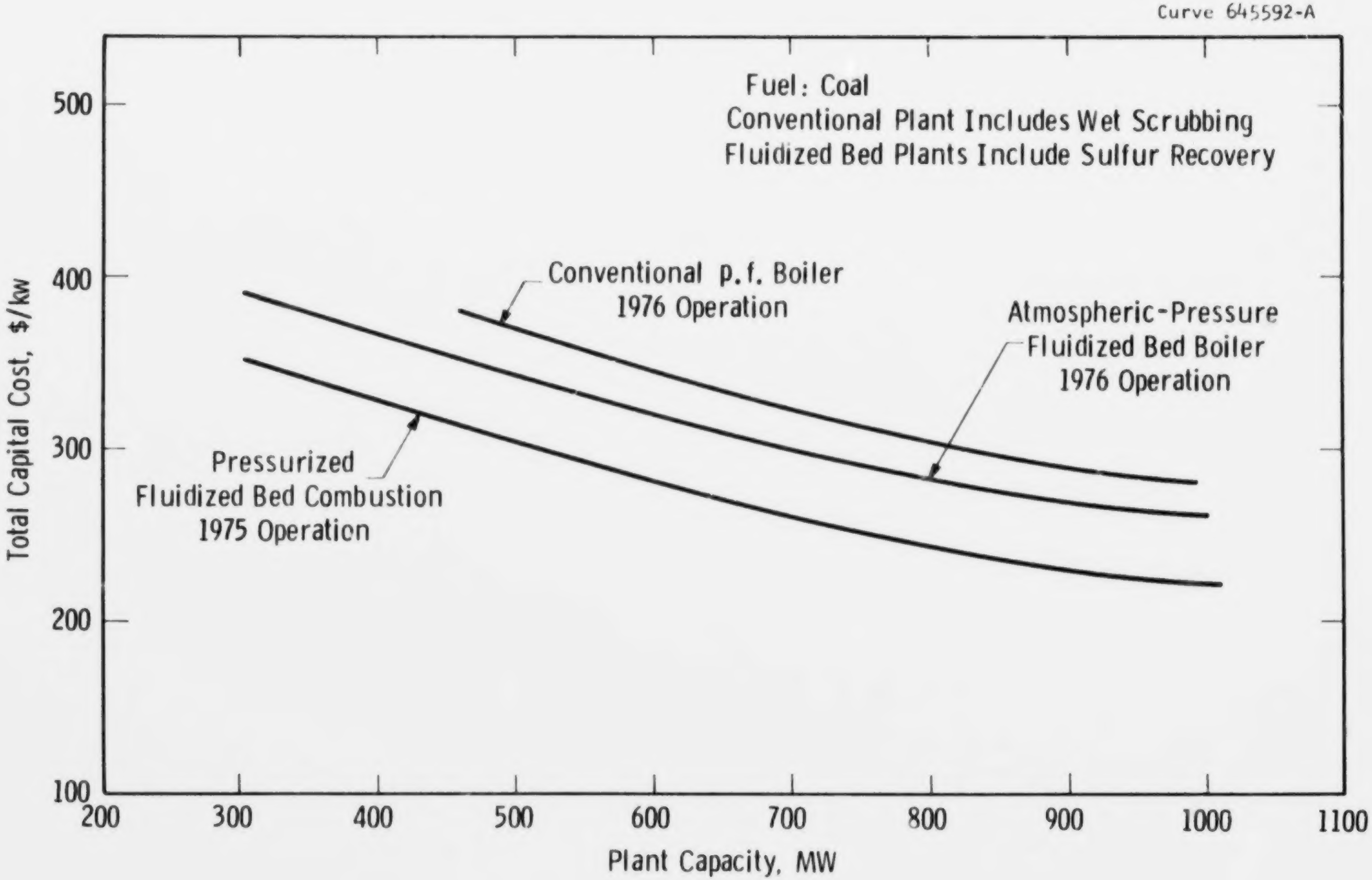

Fig. 1-153 - New Plant Cost vs. Capacity

TABLE 1-51

AIR POLLUTION CONTROL EQUIPMENT COST

	CONVENTIONAL P.F. BOILER $/KW	ATMOSPHERIC PRESSURE FLUID BED BOILER $/KW		PRESSURIZED FLUID BED BOILER $/KW	
		Regeneration	Once-Through	Regeneration	Once-Through
SO_2 removal	20[a]	(b)	(b)	(b)	(b)
Sulfur recovery	--	15.2	--	12.68	--
Particulate control[c]	(a)	5.0	5.0	(d)	(d)
NO_x minimization	3.50	(e)	(e)	--	--
	23.50	20.2	5.0	12.68	--

(a) The wet scrubber is assumed to achieve adequate particulate removal. Recent cost data [Second Annual Lime/Limestone Wet Scrubbing Symposium, New Orleans, La., November, 1971] indicate that the cost for SO_2 and particulate control by wet scrubbing for a new plant will be ∿ $40/kw

(b) Sulfur dioxide is removed in the boiler.

(c) Ash handling cost is not included.

(d) Particulate removal is required for gas turbine operation. This cost is not attributed to pollution control.

(e) Additional NO_x control may be required to meet proposed regulations.

tornado units and the gas turbine. The present design uses the high-alloy steel to assure protection of the gas turbine from additional particulates.

- Stone Regeneration System: The present design is based on a multistep process which separates each function and utilizes coal as fuel. Advanced concepts were proposed which combine two or more process steps into a single operation. If such a concept is successful, the regeneration system cost could be reduced \$3 to \$4/kw. The use of natural gas or oil as a fuel for the present regeneration system concept could result in savings of approximately \$1/kw.
- Coal Feeding System: The coal feeding system design is based on systems which have been built and operated. The design provides a separate coal feeding system for each fluidized bed in order to assure control of the coal feed rate to each bed. It may be possible, however, to reduce the number of coal feed systems from 16 to 4 if independent control of solids flow to each bed in a module can be achieved from a single pressurized injector. The potential cost reduction is estimated to be \$1 to \$2/kw.
- Stack Gas Cooler Design: Cost estimates were obtained for the stack gas coolers, but no attempt was made to optimize the design or consider nonconventional designs, such as those using fluidized beds. Preliminary conceptual evaluation indicates that the cost might be reduced $\sim$ \$1/kw from the present \$4.40/kw.

Capital cost savings may also be realized by an increase in plant performance over that projected. The gas turbine temperature of 1600°F is based on a temperature drop of 150°F of the flue gas between the fluidized bed and gas turbine. Two factors indicate this may be well below the actual gas turbine temperature: burning above the bed is expected to increase the gas temperature above the 1750°F bed temperature, and the heat losses are not expected to be as large as projected. If the gas turbine temperature were increased to 1700°F, the total capital cost reduction would be equivalent to \$6 to \$7/kw.

The total capital cost would be reduced ∿ \$20/kw if all the potential savings in equipment cost and improvement in plant performance were achieved. This would give a plant cost of ∿ \$255/kw -- about 25% less than the conventional plant cost.

For a plant using a cooling tower instead of a once-through cooling system, an additional saving in the cooling tower cost can be credited to the pressurized system over a conventional plant because of the reduced heat rejection from the condenser in a combined cycle plant and the higher plant efficiency. The cost estimate for cooling water equipment for a conventional plant is ∿ \$8/kw. The cost savings will depend on the combined cycle plant efficiency.

These design alternatives do not represent cost savings which might be achieved by advanced concepts in boiler design, sulfur recovery processes, or power cycles.

The pressurized plant could also be operated by using a once-through dolomite system instead of regenerating stone and recovering sulfur or sulfuric acid. The capital cost for this system is estimated to be \$264/kw.

Atmospheric-Pressure Fluidized Bed Boiler Plant. The preliminary design of the atmospheric pressure fluidized bed boiler was made by Foster Wheeler for a nominally 300 MW plant. The boiler cost estimates were extended to 635 MW capacity. The total capital cost estimate for the plant with sulfur recovery is \$312/kw. This represents a 7% cost reduction over a conventional plant with stack gas cleaning. The cost of a once-through limestone system is estimated to be \$295/kw.

Cost reductions are not projected for the atmospheric pressure system design. The auxiliary equipment to the boiler incorporates an advanced coal feeding system and a single-stage regeneration system. Thus, additional possible cost reductions in system components have not been identified. Advanced concepts in boiler design and advanced steam conditions may offer the potential for additional savings.

Capital costs were projected for oil-fired plants burning high-sulfur residual oil. Capital costs for 635 MW oil-fired power plants are summarized in Tables 1-52 and 1-53. Capital cost as a function of capacity is shown in Figure 1-154 Capital cost for pressurized fluidized bed combustion combined cycle oil-fired plants with sulfur recovery may be ∿ 14% less than conventional plants with wet scrubbing without sulfur recovery. The atmospheric pressure fluidized bed boiler plant is only 2% less than a conventional plant.

These capital costs are higher than many reported in other studies. However, since the costs for the various types of plant were developed on the same basis, it is thought that they provide valid comparisons.

Energy Costs

Energy costs for coal- and oil-fired plants are summarized in Tables 1-54 and 1-55. Energy cost reductions for pressurized fluidized bed combustion combined cycle coal-fired plants with sulfur recovery may be ∿ 10% or 1-1/4 to 1-1/2 mills/kwh below conventional plant energy costs. The reduction in energy costs for atmospheric pressure fluidized bed combustion with sulfur recovery is estimated to be ∿ 2%. Energy cost reductions for pressurized fluidized bed combustion combined cycle oil-fired plants with sulfur recovery may be ∿ 7% below conventional plant energy costs. Atmospheric pressure fluidized bed systems do not show significant energy cost reductions over new conventional plants.

TABLE 1-52

BOILER PLANT EQUIPMENT COSTS

	CONVENTIONAL(a)	ATMOSPHERIC PRESSURE FLUID BED COMBUSTION	PRESSURIZED FLUID BED COMBUSTION
Steam generator(b)	29.40	26.80	12.90
Draft system			
FD fans	0.90	1.50(c)	--
Precipitators	--	5.00(c)	7.50(d)
Draft flues/supports	2.63	1.66(c)	4.43(c)
Stack/foundation	0.47	0.47(c)	0.47(c)
Fue[illegible]l equipment	8.60	9.60(e)	11.00(e)
Ash handling	--	2.69(c)	0.55(c)
NO_x control	1.70(f)	--	--
Instrumentation	2.75	3.10(c)	3.10(c)
Miscellaneous	0.86	0.94(c)	0.94(c)
TOTAL	47.31	51.76	40.89

(a) Based on cost data from United Engineers and Constructors.

(b) Includes air heater cost for atmospheric pressure systems. Boiler cost for oil gasification assumed to be the same as conventional boiler--this will depend on burner cost, flame characteristics, etc. Fluid bed boilers assumed to be 10% less than coal-fired boiler designs since no carbon burn-up cell is required and the gas flow is reduced. No credit taken for reduced fouling from vanadium and sodium in fluidized bed systems.

(c) Costs same as coal-fired plant.

(d) Primary collectors in coal-fired plant not included since there is no carbon burn-up cell.

(e) Cost will depend on oil piping and pump cost to gasifier or combustor relative to conventional cost to burners. This cost is 20% of the fuel oil equipment cost in a conventional system. The cost for oil gasification is not expected to increase since the pressure and number of feed points are similar. The cost for fluid bed combustion boilers is projected to increase due to more feed points and a higher pressure requirement for the pressurized system.

(f) Flue gas recirculation.

TABLE 1-53

OIL-FIRED POWER PLANT COST BREAKDOWN
635 MW Plant

	CONVENTIONAL(a) $/KW	ATMOSPHERIC PRESSURE FLUID BED COMBUSTION $/KW Regenerative	PRESSURIZED FLUID BED COMBUSTION $/KW Regenerative
Land & land rights	1.13	1.13	1.13
Structures & improvements	24.95	25.80(b)	19.57(b)
Boiler plant equipment	47.31	51.76	40.89
Sulfur removal equipment	20.00(c)	14.10	12.68
Gas turbine-generator equipment	--	--	14.80
Steam turbine-generator equipment	55.76	54.96	44.14
Electric plant equipment	14.40	14.40	15.30(d)
Miscellaneous plant equipment	4.08	4.08	3.45(d)
Undistributed costs	28.55	27.80(d)	27.00(d)
Other plant costs	2.91	2.91	2.91
Subtotal	199.09	196.94	181.87
Normal contingency	11.95	11.82	10.91
Subtotal	211.04	208.76	192.78
Escalation	47.48	46.19	36.15
Subtotal	258.52	254.95	228.93
Interest during construction	43.63	40.14	30.05
General items and engineering	5.5	5.5	5.5
TOTAL CAPITAL COST	307.65	300.60	264.48

(a) Assumes stack gas scrubbing system, no particulate control beyond scrubber, no additional stack height over alternates. No by-pass around the scrubber system is provided.

(b) Same as coal-fired plant.

(c) Once-through throw-away system.

(d) Estimated by assuming same percent reduction for oil-fired plant as was assumed with coal-fired plant over conventional coal-fired plant.

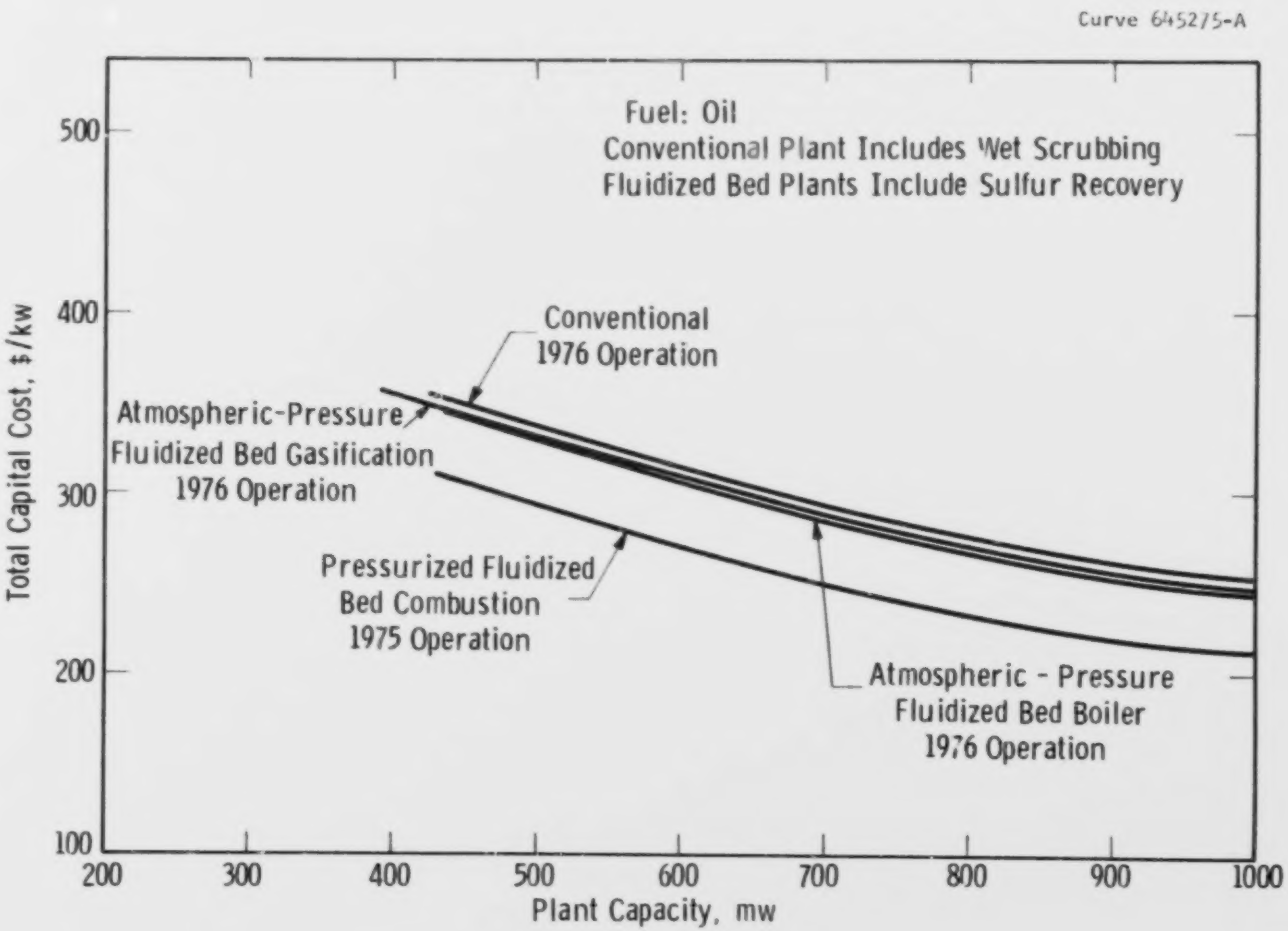

Fig. 1-154 — New plant cost vs capacity

TABLE 1-54

ENERGY GENERATION COSTS FOR COAL-FIRED POWER PLANTS
635 MW PLANT

	CONVENTIONAL PLANT	ATMOSPHERIC PRESSURE FLUIDIZED BED BOILER		PRESSURIZED FLUIDIZED BED BOILER	
		Regenerative	Once-through	Regenerative	Once-through
Fixed charges	8.10	7.61	7.20	6.75	6.44
Fuel	4.11	4.44	4.15	4.31	4.00
Dolomite or limestone	--	0.05	0.29	0.12	0.52
Operating and maintenance	0.99	0.86	0.67	0.90	0.71
Total, mills/kw h	13.20	12.96	12.31	12.08	11.67

TABLE 1-55

ENERGY GENERATION COSTS FOR OIL-FIRED POWER PLANTS
635 MW PLANT

	CONVENTIONAL BOILER (ONCE-THROUGH)	ATMOSPHERIC PRESSURE FLUIDIZED BED BOILER (REGENERATIVE)	PRESSURIZED FLUIDIZED BED BOILER COMBINED CYCLE (REGENERATIVE)
Fixed charges	7.52	7.35	6.47
Fuel	4.11	4.60	4.35
Dolomite or limestone	0.05	0.05	0.12
Operating and maintenance	0.91	0.78	0.82
TOTAL, mills/kwh	12.59	12.78	11.76

Evaluation

The fluidized bed combustion systems are compared with each other and with conventional systems with stack gas cleaning. Potential advances and uses for fluidized bed combustion boilers are reviewed.

Comparison

A comparison of atmospheric and pressurized fluidized bed combustion power systems must consider several factors:

- Design characteristics
- Effectiveness in pollution abatement
- Effectiveness in fuel utilization
- Economy of power generation
- Status of technology
- Development requirements.

Design Characteristics. A summary of the design characteristics for each system is presented in the next matrix. The heat transfer characteristics -- heat and release rates and heat transfer surface -- are compared in Table 1-56. The height of the fluidized bed boilers is approximately 100 ft for both 300 MW and 600 MW capacities. Conventional coal-fired boiler height is approximately 200 ft. The pressurized concept is modular and can, for the most part, be shop-fabricated. This greater degree of standardization and shop fabrication will allow the power industry to reduce costs and to capitalize the experience obtained during initial development.

Heat Transfer Surface. The salient advantage of the fluidized bed boiler over the conventional boiler is the reduced heat transfer surface. A comparison of the surface requirement for the atmospheric and pressurized fluidized bed boilers with a conventional p.f. boiler is as follows:

COMPARISON OF HEAT TRANSFER SURFACE AREAS

Function	Conventional Boiler(a)	Fluidized Bed Boiler	
		Atmospheric(b)	Pressurized(c)
Evaporator	2.97×10^4	5.28×10^4 (FB)	1.21×10^4 (FB)
Superheater I	8.16	1.75 (FB)	1.27 (FB)
Superheater II		1.75 (FB)	1.67 (FB)
Reheater	11.3	1.65 (FB)	1.67 (FB)
Economizer	15.36	11.44 (CONV)	1.906 (CONV)
Total	37.79	21.86×10^4	9.918
	Regenerative	Tubular	
Air Preheater	53.0×10^4	208×10^4	

(a) Georgia Power & Light Ft. Hammond No. 4, 530 MW, 2486/1000/1000, 3.6×10^6 lb. steam/hr.

(b) 600 MW, 2400/1000/1000, 3.8×10^6 lb. steam/hr.

(c) 600 MW, 2400/1000/1000, 3.26×10^6 lb. steam/hr.

Pollution Abatement Potential. The projected emissions to the atmosphere from the fluidized bed boiler plants are compared with a conventional plant with stack gas scrubbing in Table 1-57. These emissions are consistent with the plant designs and costs. Further reduction in emissions is possible for each system but would represent additional costs. The solids disposal problem is significantly reduced with the fluidized bed combustion system where the waste solids are dry and large -- 1/16 inch. Thus, the solids are easier to handle and have a greater potential for market value[25] than the slurries produced from

FLUIDIZED BED BOILER DESIGN CHARACTERISTICS

DESIGN CHARACTERISTICS	UTILITY	
	ATM. PRESSURE	PRESSURIZED
Capacity		
Power	600 MW	635 MW (17.4% gas turbine power)
Steam, lb/hr	3.8×10^6	$\sim 3.4 \times 10^6$
Conditions	2400/1000/1000	2400/1000/1000
Design Features		
Water circulation	once through forced circulation	once through forced circulation
Modules		
Number	4	4
Beds/module (incl.CBC)	6	5
Bed orientation	stacked	stacked
Heat transfer surface		
Arrangement	horizontal serpentine	horizontal serpentine
Tube size, in.	1-1/2 & 2	1-1/2 & 2
Bed depth, ft.	2.5	10-16
Gas velocity, fps	6-10	5.6-9.1
Bed temperature, °F		
Design point	1600	1750
Part load	≥ 1300	≥ 1300
CBC	1900-2000	1900-2000
Particle size, in.	-1/8 or -1/4	-1/8 or -1/4
Excess air, %		
Primary beds	10	used 10%
CBC	50	80
Total	23.6	15.8
Control method	bed temperature and module shut-down	bed temperature and module shut-down
Sulfur removal		
Stone	limestone	dolomite
Ca/S ratio	6	6
Make-up, %	~10	~10
Coal feeding		
Primary bed area, ft^2	312	70
Total bed area, ft^2	7500	1200
Feed points to each primary bed	~30	4-8

TABLE 1-56

HEAT TRANSFER CHARACTERISTICS

	POWER PLANT CONCEPT		
	Conventional p.f. Boiler	Atmospheric Pressure[a] Fluidized Bed Boiler	Pressurized[b] Fluidized Bed Boiler
Heat release rate			
Boiler cross-section			
Btu/hr - ft^2	2.2 x 10^6	0.7 x 10^6[c]	4.2 x 10^6[c]
Bed volume			
Btu/hr - ft^3	-	0.3 x 10^6	0.3 x 10^6
Boiler volume[d]			
Btu/hr - ft^3	1.7 x 10^4	5.1 x 10^4	20. x 10^4
Heat transfer surface,[e] ft^2/MW	740	410	160

[a] Based on [illegible]-Foster Wheeler 280 MW boiler design.

[b] Based on [illegible]-Foster Wheeler 317 MW boiler design.

[c] Average which includes carbon burn-up cell.

[d] Boiler volume within water walls.

[e] Excludes air preheater or stack gas coolers.

TABLE 1-57

AIR POLLUTION ABATEMENT

	POWER PLANT CONCEPT		
	Conventional[(a)]	Fluidized Bed Combustion	
		Atmospheric Pressure	Pressurized
Sulfur dioxide (lb $SO_2/10^6$ Btu)	1.5	0.7	0.7
Nitrogen oxides (lb $NO_2/10^6$ Btu)	1.2-1.4	0.4-0.8	0.1-0.2
Particulates (lb/10^6 Btu)	0.05-0.15	0.05[(b)]	0.15[(c)]

BIAS: 4.3% Sulfur, 8.5% Ash Coal

(a) With stack gas scrubber for sulfur and particulate control; no NO_x control.

(b) Electrostatic precipitator.

(c) Does not include particulate removal after the gas turbine.

wet scrubbing systems. The atmospheric pressure system also offers the advantage of large solids relative to the conventional plant. A conventional steam or water transport system is projected for solids handling, which increases the solids handling problems relative to the pressurized system.

Fuel Utilization. The plant heat rates and alternate fuel capabilities are summarized in the next matrix. The atmospheric pressure fluidized bed boiler plant has a lower heat rate than the conventional plant because of the higher carbon loss. The combined cycle pressurized fluid bed boiler plant has a higher efficiency than the conventional plant. No attempt was made in the present design to exploit the potential of the combined cycle plant to achieve higher plant efficiencies. The fluidized bed plants have the potential for burning low-grade fuels which cannot be burned in conventional plants or, if they are burned, result in high plant cost or low plant efficiency and plant availability. The low bed temperatures permit high-alkali fuels to be burned without corrosion or fouling of the boiler tubes.(1) Since many of the western coals have high alkali and high ash content, the fluidized bed concept offers an economical means of utilizing these fuels. The Igni-fluid process illustrates the potential of fluidized beds to burn low-grade fuels for power generation.[35] The fluidized bed concept also offers potential for burning solid wastes, either as the only fuel or as a supplementary fuel.

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(1) See pp. 132-156.

TABLE 1-58
FUEL UTILIZATION

	POWER PLANT CONCEPT		
	Conventional	Fluidized Bed Combustion	
		Atmospheric Pressure	Pressurized
Plant heat rate(a)	9122(b)	9230(c)	8892
Btu/kwh	(37.4%)	(37%)	(38.4%)
Alternate fuel capability	limited to low-alkali, low-ash coals; low-vanadium oil; or gas	potential for burning high- and low-grade coals, solid wastes tailings from fuel processing facilities, oils, or gas	

(a) Based on 1-1/2 in. Hg steam turbine back pressure.

(b) Hammond #4, Georgia Power & Light.

(c) This assumes a boiler efficiency of 88.6%. If the boiler efficiency is 86.1%, as projected in the design, the heat rate would be 9500 Btu/kwh.

Development Status of Fluidized Bed Combustion. The amount of time and effort devoted to atmospheric pressure fluidized bed combustion has thus far been one order of magnitude greater than that devoted to the pressurized concept. The technical accomplishments represented by the capabilities and performance of fluidized bed boiler pilot plants, however, appear comparable. The next matrix presents technical and operating data on the Pope, Evans and Robbins (PER) atmospheric system and the National Coal Board (NCB) pressurized boiler. These two pilot plants represent the most advanced experimental work on the respective fluidized bed combustion concepts.

- The size and capacity of these two systems do not differ greatly
- The pressurized boiler carried out longer runs with fewer operating difficulties at an earlier date (January 1971) than the atmospheric system (July 1971) although it was commissioned a year later
- Results on sorption of SO_2 and reduction of NO_x were more favorable in the pressurized system (see Table 1-57)
- Adequate particle removal was achieved in the pressurized system so that the gases did not erode or deposit on turbine stator blades.

Several additional factors must be considered in comparing the accomplishments of the atmospheric and pressurized systems. The PER atmospheric boiler did operate at design conditions of pressure, temperature, and velocity; and it did illustrate operation with an auxiliary unit for burning carbon carried by combustion gases from the fluidized bed and recovered by dust removal equipment. The NCB pressurized boiler operated at conditions selected for British designs but not at those for the Westinghouse-Foster Wheeler design; it contained neither an auxiliary carbon combustion system nor a sorbent regenerator.

PRESENT EXPERIMENTAL APPARATUS(a)

	ATMOSPHERIC PRESSURE PER SYSTEM	6 ATMOSPHERIC PRESSURE NCB-BCURA SYSTEM
Primary System		NCB
Components	FBM, CBC, vertical welded wall heat transfer surface in bed, coal feed systems.	Primary bed, particulate removal, pressurized coal feed system, turbine blade cascade, horizontal heat transfer surface in bed as tube bundles.
Bed area, ft^2	8.75	8
Max. coal feed rate reported, lb/hr	$\sim$900	$\sim$600
Projected equivalent capacity, kw	$\sim$1300	$\sim$900
Longest runs, hrs		350(b)
Years of operation	3-1/2	2-1/2 (Feb. 1969)
Hours of operation OAP sponsored		435 (Aug. 1970-July 1971)
Integrated System		
Components	Primary system with sorbent regenerator	---
Longest run, hrs	156(c)	
Coal feed rate, lb/hr	460	
Projected equivalent capacity, kw	700	

(a) Includes only PER and BCURA Pressurized Apparatus.

(b) The apparatus was designed for $\sim$100 hrs continuous operation. The long-term tests were thus interrupted to refill the coal feeding system and were not interrupted due to mechanical or operating problems.

(c) Because of operating problems, there were three interruptions, two of them due to coal feeder plugging.

Both atmospheric and pressurized fluidized bed power systems require further development effort. The technical problems on which further information is required are listed in the next matrix. This information can be most rapidly and economically acquired on development scale equipment.

Economy of Power Generation. Capital and energy generation costs for coal- and oil-fired fluidized bed combustion power plants and conventional power plants were presented on pages 389-403. A comparison of these cost data shows

- Pressurized fluidized bed combustion combined cycle power plants using a high-sulfur coal fuel, with sulfur recovery, may reduce energy costs 10% below conventional plants with stack gas cleaning.
- Pressurized fluidized bed combustion combined cycle plants burning high-sulfur oil, with sulfur recovery, may reduce energy costs 7% below conventional plant burning high-sulfur oil.
- Atmospheric pressure fluidized bed combustion power plant systems burning high-sulfur coal or oil may reduce energy costs less than 2%.
- Environmental control costs for SO_2, NO_x, and particulates represent 11% to 20%(1) of conventional plant equipment costs and 5% to 7%(2) of fluidized bed plant equipment costs. The cost for control of heat rejection to water systems adds ~ 4% to the conventional plant equipment cost. The cost for the fluidized bed systems is less since only 50% to 80% of the energy is produced by steam turbine generators.

Potential

Potential energy cost reductions for fluidized bed power generation systems are projected as the result of:

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(1) 20% assumes $40/kw for SO_2 and particulate control.

(2) Does not include particulate removal required for gas turbine operation.

PROBLEMS IN FLUIDIZED BED POWER SYSTEMS
FOR INVESTIGATION IN DEVELOPMENT PLANTS

ATMOSPHERIC PRESSURE	PRESSURIZED
NO_x control	Operation of Deep Beds
Is two-stage combustion necessary? Alternative techniques for control	Temperature gradients? Particle circulation characteristics Combustion efficiency SO_2 removal NO_x minimization
Evaluate particulate carryover and control equipment	Evaluate particulate carry-over and control equipment
Regeneration system	
Solids handling Sorbent activity and utilization Feasibility of producing high S content gas	Turbine blade tests
Sulfur recovery system	Regeneration system
Boiler tube materials checked	Solids handling Sorbent activity Feasibility of producing high S content gas
Heat transfer surface studies	Sulfur recovery system
Boiler tube spacing and heat transfer coefficients above bed	Boiler tube materials checked
Operation: start-up, shutdown, load follow	Heat transfer surface studies Boiler tube arrangement and heat transfer coefficients
Advanced concepts: higher steam temperatures	Operation: start-up, shutdown, load follow
	Advanced concepts: higher gas turbine temperatures, circulating beds, higher steam temperatures

NOTE. Very little information on the development plants [illegible] transferrable from one concept to the other. For example, work on large cross-section shallow beds provides no insights into the potential problems with deep beds; particulate control equipment will operate at different temperature and pressure levels; regeneration systems have to process stones with entirely different histories; the pressure drop across the distributor can be much greater in the pressurized design, but the design may be more sensitive for the large area shallow beds; and boiler tube materials are subjected to different temperatures and pressures. Work on a sulfur recovery system might be common, although even this is not yet clear.

- Modification of the present design concepts
- Advanced boiler-regeneration-sulfur recovery designs and power cycles
- Utilization of low-grade fuels.

Modification of Present Concepts. Capital cost reductions from simplifications and improvements in the present design concepts were discussed on pages 390 and 396.

The potential capital cost reduction for the pressurized system is estimated to be ~\$20/kw. This corresponds to an energy cost reduction of ~0.50 mills/kwh, a 40% reduction in the present design energy cost. If the gas turbine temperature can also be increased as the result of burning above the bed or lower heat losses than projected, the plant heat rate would increase 1% to 2%, resulting in an additional energy cost reduction of ~0.1 mills/kwh. The potential energy cost reduction over the conventional plant with wet scrubbing using the modifications in the present design with sulfur recovery is 13%. Modification in the present atmospheric pressure boiler plant design concept does not appear to result in significant cost reductions.

Advanced Concepts. Advanced designs have the potential for reducing capital costs and reducing the plant heat rate over the present design concepts. Three areas have been considered: boiler design, regeneration-sulfur recovery processes, and power cycle conditions.

- Advanced boiler design concepts which have potential advantages in maintaining uniform bed temperature, in feeding reactants to the bed, in separating ash, and in power plant control were considered. One of these is a deep, recirculating, fluidized bed boiler concept.
- Regeneration-sulfur recovery processes have also been assessed in order to develop a process with reduced capital cost and improved operability. Particular emphasis was placed on processes to eliminate a separate CO_2 generation system or for an H_2S enrichment system.

- The power cycle operating conditions can be advanced to improve the plant heat rate -- increased gas turbine temperature and increased system temperature and pressure.

In the high-pressure combined cycle system the bed temperature is limiting, rather than the gas turbine inlet temperature. If a way is found to decrease the temperature difference between the bed and the gas turbine, or if a supply of gaseous fuel can be obtained for heating the exhaust products from the beds to the current limiting level for the gas turbine, i.e., 1800°F-1900°F, the plant heat rate would decrease 2% to 3%, and the plant capacity would increase 5% to 8%. The effect on specific plant cost would be somewhat less than the capacity change. Gas turbine temperature of 2500°F is projected for 10 to 15 years from now, so the potential for growth in this area is substantial. No significant improvements are expected in gas turbine aerodynamic efficiencies in the foreseeable future. There is, however, a definite trend in gas turbine design toward higher mass flows on an area basis. This will bring about further reduction in specific plant cost. Reduced corrosion in fluidized bed boilers could result in increasing steam temperatures up to 1200°F or higher. Both the plant capacity and the plant heat rate would change about 4% per 100°F increase. Near-term improvements in gas turbine and steam turbine utilization could result in heat rate improvements on the order of 10% and in specific plant cost reduction on the order of 10% to 15%. Longer term advances in gas turbine temperatures would bring about further improvements.

The potential energy cost reductions for the atmospheric pressure and pressurized fluidized bed boiler plants based on successful development of the advanced concepts are summarized below. The energy costs for the atmospheric pressure and pressurized systems represent reductions of 11% and 23%, respectively, over a conventional plant with stack gas cleaning:

TABLE 1-59

ADVANCED FLUIDIZED BED COMBUSTION POWER PLANT COSTS
WITH SULFUR RECOVERY
BASIS: 600 MW

	ATMOSPHERIC PRESSURE PLANT	PRESSURIZED PLANT
Capital cost, $/kw	290	240
Heat rate, Btu/kwh	8300	7600
Energy cost, mills/kwh	11.90	10.30

Fuel Utilization. Fluidized bed combustion boilers offer a means for burning low-grade fuels which cannot be effectively burned in conventional boilers. Two advantages can be realized: effective utilization of fuel resources and reduction in energy cost as the result of low-cost fuel supplies.

Fluidized bed combustion boilers may also be attractive for burning char produced from coal gasification for power generation. A possible combination of gasification and combustion is shown in Figures 1-155 and 1-156. The gasifier produces a clean fuel gas by driving off the volatile components from the coal and removing sulfur. The residual devolatilized char from the gasifier is burned in a fluidized bed combustion boiler which also acts as a heat recovery boiler for the gas turbine exhaust. The system may have improved economics and provide improved turn-down capability by storing char.

Assessment. In comparing technology, the state of development and the economics of the problem must be considered and a balanced assessment rendered. Demonstration installations of wet scrubbers are well underway; 70% to 90% removal of SO_2 can be achieved; some operating problems -- corrosion and deposition -- must be solved. Costs of new scrubber installations -- about $20 to $40/kw -- have been estimated with fair accuracy. Combustion modification techniques on large conventional boilers have been under investigation and it appears that NO_x emissions can be reduced.

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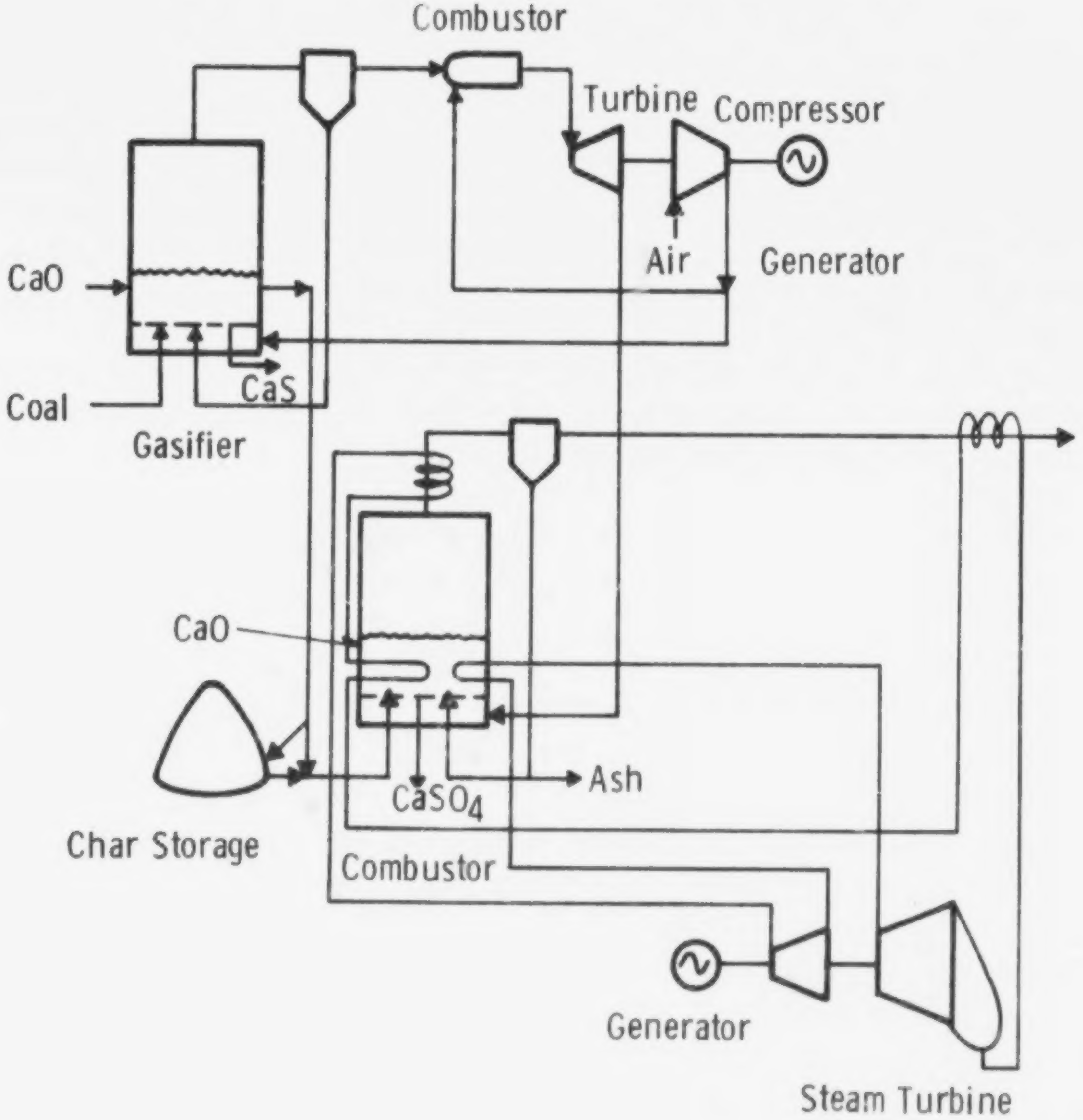
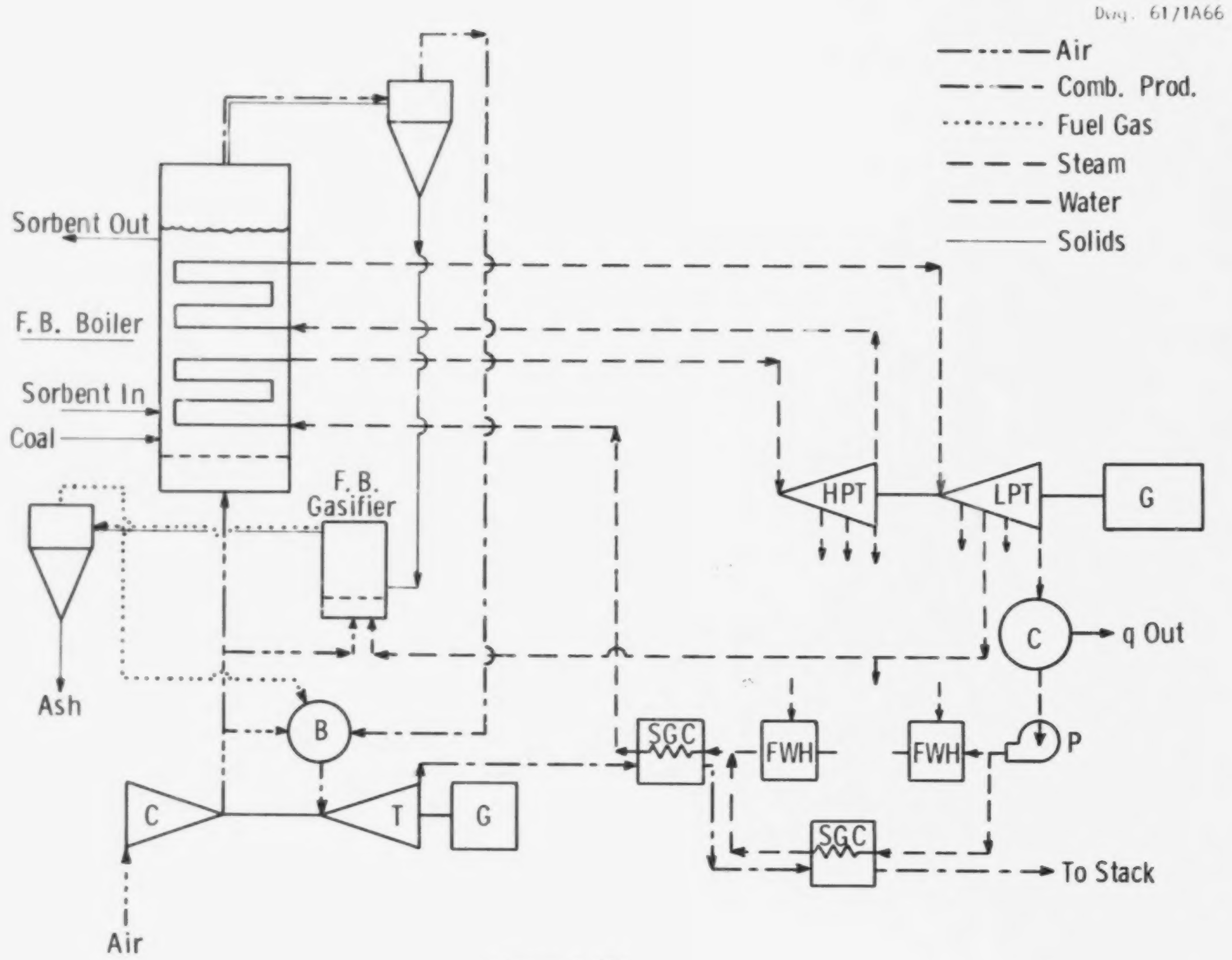


Fig. 1-155 –Fluidized bed gasification-combustion power plant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Air
Comb. Prod.
Fuel Gas
Steam
Water
Solids
Sorbent Out
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Sorbent In
Coal
F. B. Gasifier
Ash
B
C
T
G
Air
HPT
LPT
G
C
q Out
P
SGC
FWH
FWH
SGC
To Stack

Figure 1-156

Atmospheric fluidized bed combustion is still in a developmental stage. Demonstrated SO_2 removal is 80% to 90%; NO_x reduction may not be adequate to meet emission standards. The technology of sorbent regeneration and sulfur recovery is not sufficiently developed for accurate systems design and economic evaluation. With reasonably optimistic assumptions, energy costs from atmospheric fluidized bed boiler power plants are estimated to be 2% to 3% less than those from conventional plants equipped with wet scrubbers. This cost advantage might well disappear as further technical data are developed on the atmospheric system.

Pressurized fluidized bed combustion is also in the development stage. Demonstrated SO_2 and NO_x reductions are adequate to meet emission standards. Energy costs are projected to be 10% and potentially 24% less than conventional power plants. There is thus a much larger economic margin for solving technological problems.

System evaluation and experimental test results indicate that the pressurized system is more attractive for pollution abatement and for economy of operation. A summary of the plant comparisons and potential is presented in Table 1-60. The commercialization of either the atmospheric or the pressurized system would require 9 to 11 years and $\$65$-$75 \times 10^6$, primarily because the construction of a demonstration plant dominates the time and cost estimates.

The number of common technical problems remaining in atmospheric and pressurized systems appears very small (see page 414). The order of magnitude differences in pressure, in mass flow per unit of bed area, and in bed depth make it difficult to transfer technical information and design procedures.

In conclusion, the conventional power plant with combustion modification for NO_x control and wet scrubbers for SO_2 and particulate control is the most highly developed technology. Pressurized fluidized bed combustion technology is the most promising of its kind from the point of view of effective pollution abatement, efficient fuel

TABLE 1-60

COAL-FIRED POWER PLANTS
BASIS: 600 CAPACITY
4.3% S COAL

		FLUIDIZED BED			
	CONVENTIONAL WITH SCRUBBING SYSTEM	ATM. PRESS.		PRESSURIZED	
		REGENERATION	ONCE-THROUGH	REGENERATION	ONCE-THROUGH
Cost					
Installed capital cost, $/kw	337	313	295	277	264
Power cost, mills/kwh w/o sulfur credit	13.40	13.12	12.46	12.12	11.71
Environmental Factors					
Plant heat rate (exclusive of sulfur recovery or removal systems) Btu/kwh	9230	9550		8967	
SO_2, (lb/10^6 Btu)	1.5	0.70		0.70	
NO_x, (lb NO_2/10^6 Btu)	1.2-1.4 w/o recir.	0.4-0.8		0.1-0.2	
Particulate, (lb/10^6	0.05-0.15	0.05		0.05-0.15	
Potential for Advanced Concepts					
Cost - capital	~cost projected above	285-290		240-250	
Cost - energy		11.90		10.10-10.50	
Efficiency	No increase projected	40%		47%	
Fluidized Bed Boiler Systems					
Relative advantages (technical)		Less stringent particulate removal More exp. data on operation		Coal feeding NO_x removal Control Higher efficiency Reduced size Reduced heat transfer surface Greater potential Reduced construction time	
Technical problem areas		Coal feeding NO_x emission Control SO_2 removal		Deep beds Gas turbine blade erosion SO_2 removal Solids handling	

utilization, and economical power generation. Atmospheric fluidized bed combustion may be better than conventional systems, but further development is required.

2. ASSESSMENT OF A FLUIDIZED BED OIL GASIFICATION-COMBUSTION POWER SYSTEM

INTRODUCTION

Atmospheric pressure residual oil gasification/desulfurization concepts are being studied under contract with OAP by Westinghouse and Esso (England) to make possible the on-site production of a low-sulfur fuel gas suitable for power plant utilization. A number of factors indicate a promising market potential for such a process: stricter SO_2 emissions standards, unstable prices for low-sulfur fuels (both coal and oil in the 0.3 to 0.5 percent sulfur range), and a growth in the use of residual oil for power generation on the East coast of the United States, in both new plants and conversions, (expected annual growth rate between 1969 and 1975, 13.5 percent).

Fundamentally, oil gasification/desulfurization consists of a complex cracking-partial combustion phenomenon in a fluidized bed (the gasifier) of lime. H_2S evolved during this complex operation is simultaneously absorbed by the lime bed to yield a clean fuel gas and a sulfided lime at a control temperature of 1600°F. The hot, low-sulfur, fuel gas from the gasifier is transported directly to the boiler, where combustion is completed. Two operational modes for this process are considered in this preliminary study: the regenerative mode and the once-through mode. In the regenerative mode, the utilized gasifier lime is continuously circulated to an air-fluidized bed regenerator in which an SO_2-rich gas stream is produced. This SO_2 stream is converted to elemental sulfur, or sulfuric acid, in a sulfur recovery section of the system, while the regenerated lime is returned to the gasifier along with a nearly stoichiometric amount of fresh make-up limestone. On the other hand, the once-through mode utilizes the gasifier lime to a much greater extent than does the regenerative mode, and the sulfided lime is disposed of directly, leading to a stone disposal rate three to four times that of the regenerative mode; but the overall operation is simpler. This study investigates both modes to evaluate quantitatively the merits and disadvantages of each.

Esso (England) has provided experimental information on the oil gasification/desulfurization process based on small-scale batch fluidized units (seven-inch diameter beds). Concurrently with the batch experimental program, Esso (England) has constructed an approximately 1 MW continuous gasification/desulfurization system (operated regeneratively) which is now in its commissioning run. This batch experimental work has provided the basis for preliminary design considerations.

The scope of the preliminary investigation includes:

- The evaluation and scale-up of the Esso (England) batch data
- The generation of conceptual system designs
- The evaluation of both new boiler concepts and retrofit boiler concepts in terms of feasibility and performance
- The development of preliminary cost estimates and overall system evaluations
- Recommendations for future design considerations and further experimental studies to be carried out by Esso (England) (both continuous and batch)
- Recommendations on the development of a utility demonstration plant.

In general, the investigation considers boiler capacities greater than 100 MW, with interest centered on high sulfur removals (90% to 95%).

The basic operating variables are identified from the Esso (England) nonsteady batch studies for both modes of operation, and this information is interpreted in terms of the large-scale continuous operations of a commercial system (> 100 MW). The data indicate a potential for 90% to 95% sulfur removal in a commercial unit under a variety of operating conditions for both regenerative and once-through operations. Information to be obtained from the Esso (England) continuous unit will reveal more scale-up information needed in a detailed design phase.

Material and energy balances are computed for the conceptual system designs proposed with once-through and regenerative operations. The methods used to maintain temperature control in the gasifier, regenerator, and sulfate generator are design variables of major importance in those computations, and the results provide both definite conclusions and an indication of the need for further experimental information in the area of temperature control.

The feasibility and performance of retrofit design concepts are considered on the basis of the boiler modifications needed for conversion and the performance of the retrofit boiler. Retrofit concepts are divided into two groups: "internal designs," in which the gasifier system is placed directly beneath the boiler and the burners are placed in the upshot position at the base of the boiler to provide intimate contact between the burners and gasifier; and "external designs," in which the burners are left in their normal boiler locations and the gasifier system is placed external to the boiler. Preliminary quantitative evaluations of these retrofit design concepts are made for the retrofit cases of coal- and oil-fired boilers in terms of necessary modifications and boiler performance. New boiler designs are also considered in terms of boiler-gasifier configuration. Burner design and system operation are discussed with reference to both new and retrofit boilers.

Capital and operating costs are estimated for regenerative and once-through operations, both new and retrofit boilers, for capacities of 600 MW with 80% and 40% load factors and 90% to 95% sulfur removal. The estimates point out the most important system cost contributions.

Regenerative and once-through operations are compared with each other, and both are compared with the SO_2 control schemes of low-sulfur oil and stack gas cleaning. Other environmental factors, such as NO_x, particulates, and solid waste are included in the comparison.

The preliminary analysis leads to logical design conclusions and future design recommendations, and points out some specific areas needing further experimental analysis to assist in a detailed design phase. The most advantageous characteristics of a gasification/desulfurization demonstration plant and the timing needed in the next phases of development are also stated.

GASIFICATION/DESULFURIZATION CONCEPTS

Two possible modes for gasification/desulfurization operation are the regenerative mode and the once-through mode. Figure 2-1 illustrates the major process streams and identifies the basic elements of the two operational modes, without reference to the specific system configuration or specific retrofit concepts. The operating variables and temperature control schemes differ slightly for the two modes.

The basic elements of the regenerative operation are the gasifier vessel and the regenerator vessel, as shown in Figure 2-1. In the regenerative operation, residual oil is injected into the gasifier vessel, an air-fluidized bed of lime at 1600°F operated with substoichiometric air (∿ 20% of stoichiometric), to yield by cracking, partial combustion, and H_2S absorption by the lime, a hot, low-sulfur, fuel gas and sulfided lime. The fuel gas is transported to the boiler burners, where combustion is completed, and the sulfided lime is sent to the regenerator. The regenerator is an air-fluidized vessel operated with a slight excess of air at 1900°F. Regeneration takes place by reaction of oxygen with the utilized lime to give an SO_2-rich stream (of about 10 mole % SO_2) and a regenerated lime with a decreased activity compared to that of fresh lime. The SO_2 stream is transported to a sulfur recovery system, and the regenerated lime is returned to the gasifier along with a nearly stoichiometric amount of fresh make-up limestone.

A variety of temperature control schemes may be useful in controlling the temperatures of the gasifier and the regenerator. The temperature control scheme used in the gasifier must conserve the energy of the fuel in a form usable in the boiler. For this reason water injection, steam injection, and stack gas recycle are considered for temperature control of the gasifier by the storage of sensible heat; while the use of heat transfer surface in the gasifier to preheat

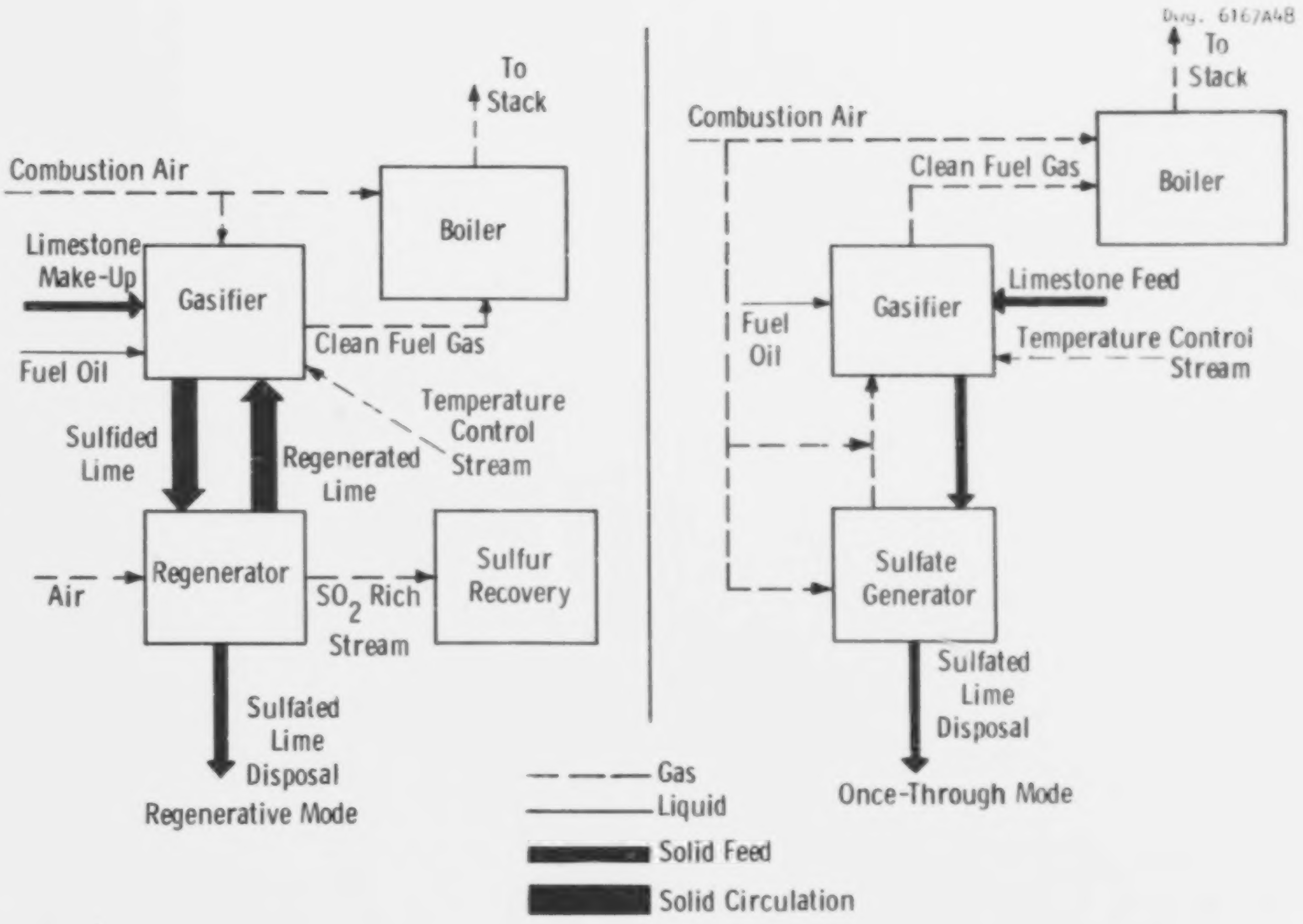

Fig. 2-1—Modes of operation

boiler water is also an attractive possibility. A fifth possibility is to reduce the air/fuel ratio to a value low enough (∿ 14% of stoichiometric) that the gasifier is thermally balanced and no further temperature control need be considered. These five possible control schemes are discussed in greater detail below.

Because of the high cost of sulfur recovery, temperature control of the regenerator is best accomplished in such a way as to avoid diluting the SO_2 stream produced by regeneration. For this reason temperature control by heat transfer surface in the regenerator, by the addition of fresh make-up stone to the regenerator, or by controlling the rate of lime circulation between the gasifier and regenerator are investigated.

The important operating variables in the regenerative operation are the air/fuel ratio, the gasifier and regenerator temperatures, the limestone particle sizes, the gasifier and regenerator fluidization velocities and bed depths, the stone residence times in both the gasifier and regenerator, and the limestone make-up rate. The effects of these variables are interrelated with the temperature control schemes used and the fuel and limestone properties.

An important phenomenon which takes place in the gasifier during the cracking-partial oxidation is the deposition of carbon, or coking, on the gasifier lime. Carbon deposition affects the sulfur removal efficiency, the thermal efficiency of the system, and the overall operation of the gasifier and regenerator when in the regenerative mode. Minimizing the rate of carbon deposition is, therefore, an important design consideration.

The basic elements of the once-through operation, shown in Figure 2-1 are a gasifier vessel and a sulfate generator, or predisposal vessel. The operation of the once-through gasifier is the same as that of the regenerative operation gasifier, while the sulfate generator operates similarly to the regenerator but at a lower temperature (∿ 1500°F). Thus, the sulfated lime from the gasifier is converted to calcium sulfate rather than calcium oxide. The calcium sulfate may be disposed of while the gas stream from the sulfate generator is sent to the gasifier. The once-through operation requires a limestone addition rate three to four times that used in the regenerative operation.

The same five temperature control methods proposed for the regenerative gasifier are considered for the once-through gasifier, while the sulfate generator temperature is controlled by one of two proposed schemes: heat transfer surface or excess air circulation.

The same operating variables are important in the once-through operation as in the regenerative operation, and carbon deposition is again a major factor in determining the sulfur removal efficiency and the overall system operation. However, in the once-through operation the carbon deposition rate does not affect the system thermal efficiency because the thermal energy of the deposited carbon is recycled as sensible heat from the sulfate generator back to the gasifier.

The two conceptual modes of operation described illustrate the basic characteristics of the systems considered in preliminary design investigations.

DATA EVALUATION

Results of the Esso (England) experimental program have identified the important operating variables and the critical phenomena associated with atmospheric pressure gasification/desulfurization. Mechanisms involved in gasification, desulfurization, and regeneration are discussed briefly, along with the essential experimental findings. The application of the Esso (England) results to the design of a utility gasification/desulfurization system is considered, and the areas of particle carryover, metal retention, burner design, and the Esso (England) continuous unit are included.

The Esso (England) Program

Under contract with OAP, Esso (England) is conducting laboratory tests on atmospheric pressure batch fluidized bed equipment to investigate lime sulfur absorption and lime regeneration operating variables [20, 66]. The batch work is being carried out in conjunction with the construction and operation of a 750 kw continuous pilot unit. The unit will be operated for 200 hours under the present contract. The information obtained from the batch tests and the continuous unit provides the basis for the design of a gasification/desulfurization retrofit system to be operated as a demonstration unit on a utility boiler.

Investigations were carried out in four different batch units, all very similar in design. All four beds were seven inches in diameter; three of the units were 5-1/4 feet high from distributor to lid, and a fourth unit was 12-1/2 feet high. Other design differences in the four units involved location of cyclones and differences in insulation on the units. Thus, results from the units are comparable and have been combined.

Two types of experimental operations have been carried out:

- Fresh bed tests, in which the bed is loaded with a given amount of fresh limestone, heated and calcined, and then operated as a gasifier-desulfurizer to note the effects of the operating variables on the sulfur removal efficiency as a function of time. After two to three hours of operation the fuel oil is no longer fed to the bed, and regeneration of the sulfided stone is recorded as a function of time.

- Cyclic tests have been carried out to investigate the continuous behavior to be expected from a regenerative process. In these tests, the procedure carried out in the fresh bed tests is repeated in cycles of gasification steps and regeneration steps with a set amount of fresh limestone added to the bed at the end of each cycle to make up for losses due to attrition and sampling, and to maintain a level of bed activity.

Two residual fuel oils have been involved in the batch tests - a U.K. fuel oil with 3.03 wt % sulfur and a U.S. fuel oil containing 2.22 wt % sulfur. Some of the fuel properties are listed in Table 2-1:

TABLE 2-1

FUEL OIL PROPERTIES

PROPERTY	U.K. FUEL	U.S. FUEL
Specific gravity	0.95	0.95
Kinematic viscosity	98.2 (140°F)	54.7 (180°F)
Carbon (wt %)	84.8	85.9
H_2 (wt %)	11.6	11.5
Sulfur (wt %)	3.03	2.22
Vanadium (ppm)	140	270
Nickel (ppm)	20	---
Sodium (ppm)	37	13
Iron (ppm)	1	4
Asphaltenes (wt %)	3.72	6.3
Conradson carbon (wt %)	9.32	7.5

From these properties, the heating values of the two fuel oils were estimated to be about 17,700 Btu/lb (low heating value). Batch operations with these two fuels were found to be nearly identical.

Three limestones have been utilized in the batch tests, and two of them, a U.K. and a U.S. limestone, have been found to function well in the gasification/desulfurization operation. Compositions of the two limestones are indicated in Table 2-2:

TABLE 2-2

LIMESTONE COMPOSITIONS

COMPONENT	U.K. LIMESTONE (wt %)	U.S. LIMESTONE (1961) (wt %)
CaO	55.2	45.0
MgO	0.27	2.02
SiO_2	0.32	14.2
Fe_2O_3	0.07	0.53
Al_2O_3	0.06	1.14
Na_2O	0.03	0.17
CO_2 released	43.4	35.4
S as SO_3	---	1.0

Esso (England) examined gasification, desulfurization, and regeneration performance as functions of the following operating variables: gasifier and regenerator temperature, air/fuel ratio, fluidization velocities, bed heights, limestone particle size, and limestone make-up rate. In applying the results of the Esso (England) batch tests to the design of a continuous commercial power plant, the following factors must be considered:

- The data are nonsteady state, batch data.
- Experiments were limited by heat losses, because temperature control could not be maintained under all conditions due to the heat losses from the small batch units. Bed heights were also limited by this factor.

- The quality of fluidization may differ in the batch units from that expected in a commercial unit; the batch units were often operated in a state of slugging, or vigorous bubbling.

Gasification

The gasification phenomenon involves the cracking operation, which leads to carbon deposition and the evolution of low molecular weight hydrocarbons, combined with the combustion of both deposited carbon and fuel to provide the heat needed to maintain the system temperature. Since the bed is operated with excess fuel, the region of the bed above the distributor and below the point of fuel injection is likely to be essentially an oxidizing region, while the zone above the fuel injection point is essentially reducing. Cracking takes place substantially in the area of the fuel injection point, while deposited carbon combustion and fuel combustion are dominant in the oxidizing zone. Gasifier product equilibrium compositions, if equilibrium does exist, are established in the reducing zone of the bed. Carbon deposition and product compositions have been studied in the Esso (England) program to provide information about the gasification operation.

Carbon Deposition

Since carbon deposition affects the overall gasification/desulfurization operation, some understanding of the phenomenon is necessary. The operating and design variables affecting the carbon deposition rate in the Esso (England) batch tests are:

- Gasifier temperature
- Air/fuel ratio
- Residual oil composition
- Fuel distribution
- Particle sizes or surface area in bed.

Lowering the bed temperature increases the carbon deposition on the lime; a lower air/fuel ratio has the same effect. Since temperature control of the gasifier was not maintained, lower air/fuel ratios led to lower bed temperatures, making the two variables interdependent. The carbon deposition should be proportional to the amount of Conradson carbon in the fuel (Conradson carbon is a measure of the amount of residue left after fractionation of the fuel). The location and quality of fuel distribution in the bed are important and are highly dependent on the quality of fluidization and solids mixing. The particle size or lime surface area in the bed is probably a factor in the deposition of carbon, but it has not been investigated directly.

If the rate at which carbon is laid down on the lime bed is assumed to be a constant, K_1, for a given fuel, stone, and air/fuel ratio, and if the rate at which carbon is burned off in the oxidizing region of the bed is assumed to be first order in the fraction of carbon on the stone, X_c, and first order in the oxygen concentration in the oxidizing region, then for perfect mixing of particles in the bed

$$W_B \frac{dX_c}{dt} = K_1 - K_2 X_c \quad , \qquad (2\text{-}1)$$

where W_B is the weight of the bed and K_2 is the product of a kinetic constant, the volume of the oxidizing zone, and the oxygen concentration. Then, for a batch test, the carbon fraction and carbon deposition rate are given by

$$X_c = K_1/K_2 \left[1 - \exp\left(-\frac{K_2}{W_B} t\right)\right]$$

and

$$W_B \frac{dX_c}{dt} = K_1 \exp\left(-\frac{K_2}{W_B} t\right) \quad ,$$

where t is the time variable. On the other hand, for a continuous operation in which lime is being withdrawn from the gasifier at a mass rate N_G, pound per pound of fuel, the carbon deposition is represented by

$$X_c N_G = \frac{K_1 N_G}{N_G + K_2} . \qquad (2-2)$$

Therefore, in a continuous operation, the carbon deposition rate will be largely dependent on the lime residence time in the bed. For short residence times ($N_G >> K_2$) the carbon deposition rate will be a constant equal to K_1 (K_1 equals the initial carbon deposition rate in a batch operation), while for large residence times ($N_G << K_2$) X_c becomes a constant equal to the equilibrium value found in a batch operation. It is felt that the latter will generally be the case with the conditions considered in this investigation.

Figure 2-2 summarizes the carbon deposition rates obtained in batch tests as a function of the air/fuel ratio and indicates the experimental conditions. Although lower air/fuel ratios necessarily meant lower bed temperatures, the data do indicate a reasonable range in which the carbon deposition should fall in a continuous operation. Also, because of the superior temperature control expected in a commercial unit, lower carbon deposition rates are expected.

Gasifier Product Analysis

Table 2-3 shows the results of analysis of the gasifier product in three of the batch runs. The chemical analysis is incomplete in that it excludes condensible tars, water, and H_2S, all of which are present in the product. The complete analysis can be estimated by using the fact that the inert nitrogen should all be present in the analyzed product gas. The material balance is completed for run 53 to yield the results shown in Table 2-4. This approximate product analysis corresponds to a hot gas thermal efficiency of 92% and a cold gas thermal efficiency of about 80%. The hot gas heating value of the product is 299 Btu/S ft^3, 12% of which is sensible heat. The total heat generated in this run by vaporization, cracking, deposition, oxidation, and sulfur removal was about 3600 Btu/hr/lb fuel/hr, 40% of which was lost to the surroundings to yield the observed gasifier temperature. Material losses in the form of tars and soot have been neglected in the product analysis, so the results are probably optimistic.

Curve 643909-A

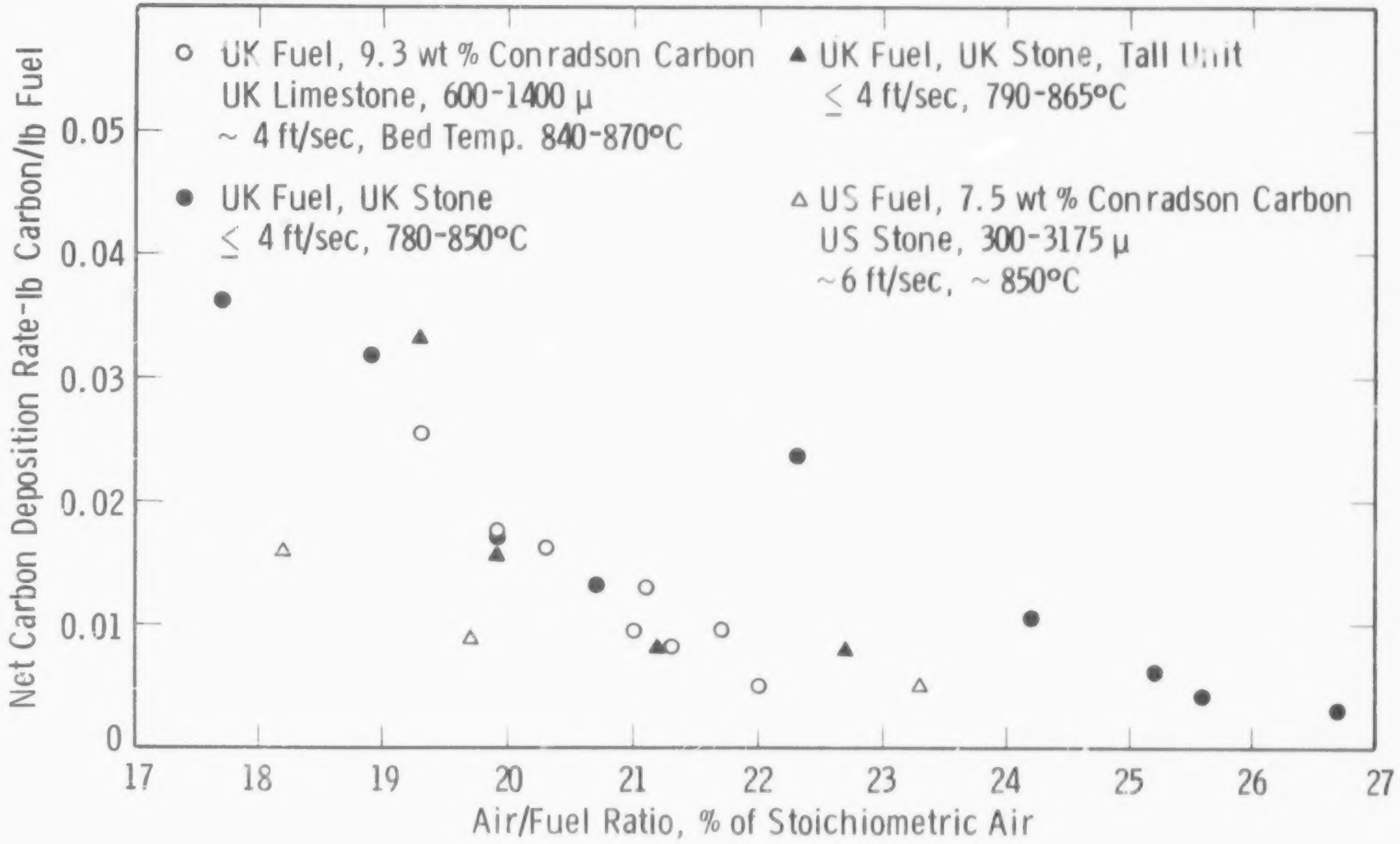

Fig. 2-2—Carbon Deposition Data

TABLE 2-3

PRODUCT ANALYSIS

	RUN		
CONDITIONS	53	65	86
Air/fuel (% of stoichiometric)	23.3	23.3	24.1
Fuel-limestone	UK-US	UK-US (1691)	UK-US (1690)
% Sulfur removed	72.5	70.3	79.1
Temperature	880°C	885°C	865°C
Component (mole %)			
N_2	64.0	57.0	62.0
CO	10.4	10.3	8.3
CO_2	10.1	9.7	10.9
H_2	4.2	10.0	7.4
CH_4	4.4	6.5	5.1
C_2	5.4	5.8	5.4
C_3	0.3	0.5	0.8
C_4	0.2	0.2	0.1

TABLE 2-4

ESTIMATED CHEMICAL ANALYSIS

COMPONENT	Wt. %
N_2	58.70
CO	9.55
CO_2	14.60
H_2	0.28
CH_4	2.31
C_2H_4	4.96
C_3H_6	0.41
C_4H_8	0.37
Condensable tars, soot	5.61 carbon + 0.92 H_2
H_2S	0.20
H_2O	1.99

The total heat production in the gasifier as a function of the air/fuel ratio can be established by carrying out the same analysis for runs 65 and 86, in addition to other data supplied by Esso giving CO and CO_2 concentrations from the gasifier. The results of this procedure, shown in Figure 2-3, are needed to compute the temperature control requirements in the gasifier.

The data also indicate a mean molecular weight of the gasifier product of 29, and a mean heat capacity of 0.35 Btu/16°F, both nearly independent of the air/fuel ratio.

The product analysis shown in Table 2-4 may be a reasonable analysis to expect from a continuous gasifier-desulfurizer operating with a similar air/fuel ratio and sulfur removal, and with no special provisions for temperature control. At other air/fuel ratios, with higher sulfur removals, and with temperature control of the gasifier by the addition of some component such as stack gas or steam, the results in Table 2-4 are not of quantitative importance. However, the design work can be carried out without complete knowledge of the product composition, since Figure 2-3 provides sufficient information on which to base the preliminary design.

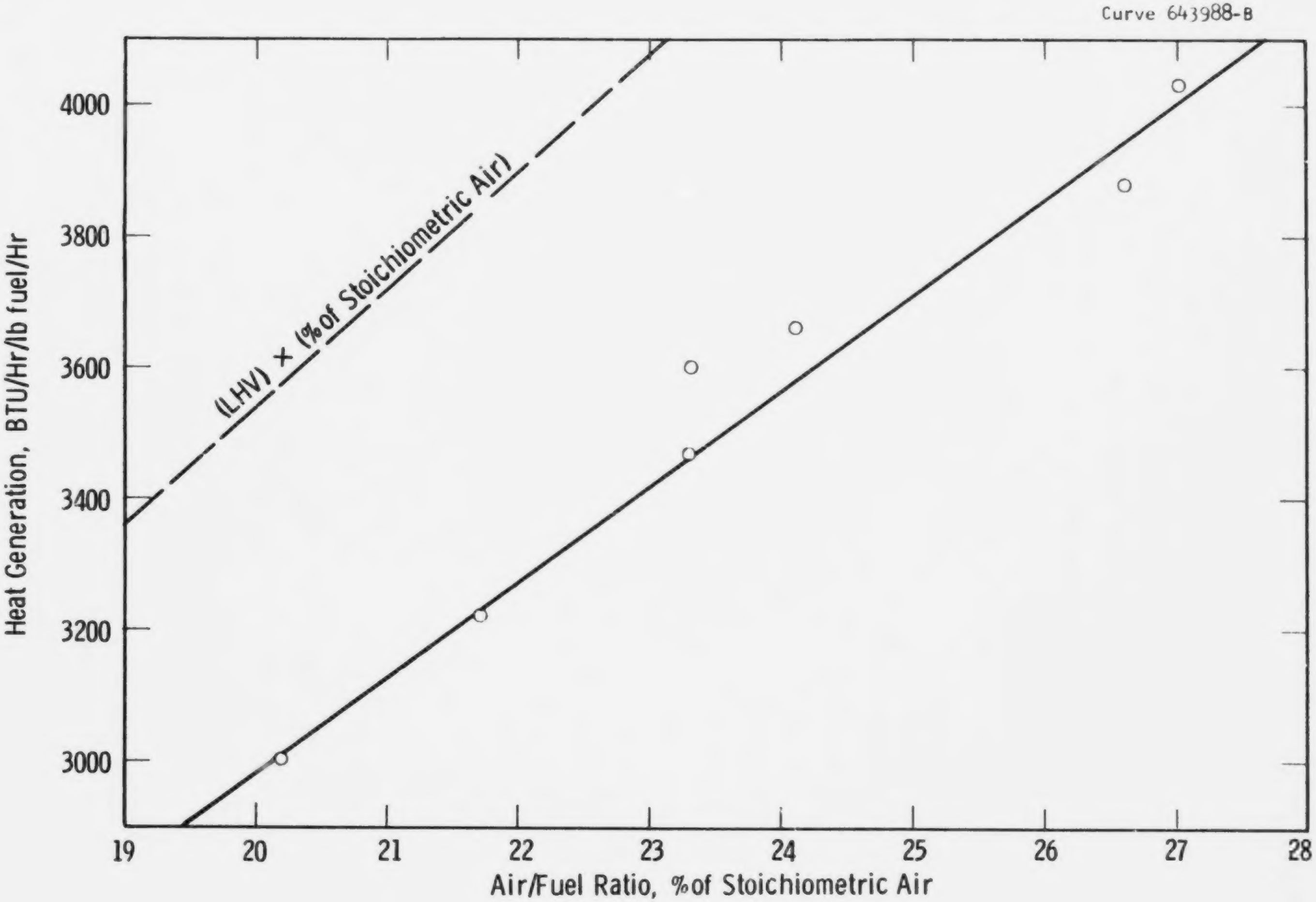

Fig. 2-3-Gasification-Desulfurization Heat Generation Rate

Desulfurization

H_2S evolved during fuel cracking reacts with the CaO in the bed by the reaction

$$CaO + H_2S \rightleftarrows CaS + H_2O \; . \qquad (2\text{-}3)$$

For both fresh bed tests and cyclic tests, the extent to which this reaction proceeds depends on the gasifier temperature, the air/fuel ratio, the gas residence time, the lime particle sizes, the bed lime content relative to the rate of sulfur fed to the bed, the amount of lime utilized, the limestone make-up rate, and the quality of fluidization. Regenerated lime loses activity compared to that of fresh lime. This may be due to a continuous deterioration in the pore structure of the lime particles brought about by the high temperatures in regeneration. Sulfate formation taking place in the regenerator may become a barrier to further sulfur removal; and carbon deposited on the lime is also believed to be detrimental to sulfur removal.

The next matrix summarizes the experimental results for sulfur removal found in atmospheric pressure fresh bed and cyclic batch tests.

For fresh bed tests operated at a temperature and air/fuel ratio at which carbon deposition does not reduce sulfur removal (temperature greater than 830°C and air/fuel ratio greater than 20% of stoichiometric), the important variables are the gas residence time, bed CaO content per pound of sulfur fed per hour to the bed, the average particle diameter, and the CaO utilization (the fraction of calcium oxide converted to calcium sulfide, or sulfate).

Figure 2-4 illustrates the results of three fresh bed tests under identical conditions of gas residence time (∿ 0.22 sec), bed CaO content per pound of sulfur fed per hour (∿ 24 lb CaO/lb S/hr), bed

SULFUR REMOVAL - EXPERIMENTAL OBSERVATIONS

VARIABLE	RANGE	COMMENTS
Gasifier Temperature	790-950°C	For best sulfur removal 830-880°C < 830°C carbon deposition increases > 880°C release of SO_2 in bed increases Temperature not independent of air/fuel rated in batch tests
Air/Fuel	15-30% or Stoichiometric	For best sulfur removal found for > 20% Air/fuel controls temperature and carbon deposition rate in batch tests carried out
Gas Residence Time	Bed depth 10-20 inches Velocity 3-8 ft/sec	Increased residence time increases sulfur removal Behavior dependent on particle size See Figure 2-5
Average Particle Diameter	4.96μ, 682μ, 1700μ	Finer particles give greater sulfur removal for a given lime utilization, bed depth, etc. See Figures 2-4 and 2-5
Bed CaO Content Sulfur Feed Rate		Increase gives greater sulfur removal Not varied independent of air/fuel ratio, velocity, bed height, etc., in batch tests See Figure 2-5
Limestone Makeup Rate	Up to 1 mole CaO per mole S fed	Nearly 100% sulfur removal obtained in cyclic tests with 1 mole CaO/mole S fed make-up rate See Figure 2-6
Quality of Fluidization		No observable effects due to increased pressure drop across distributor of higher velocities Batch tests are in a state of vigorous bubbling Good mixing must be maintained to prevent carbon build-up on bed line

Curve 643911-A

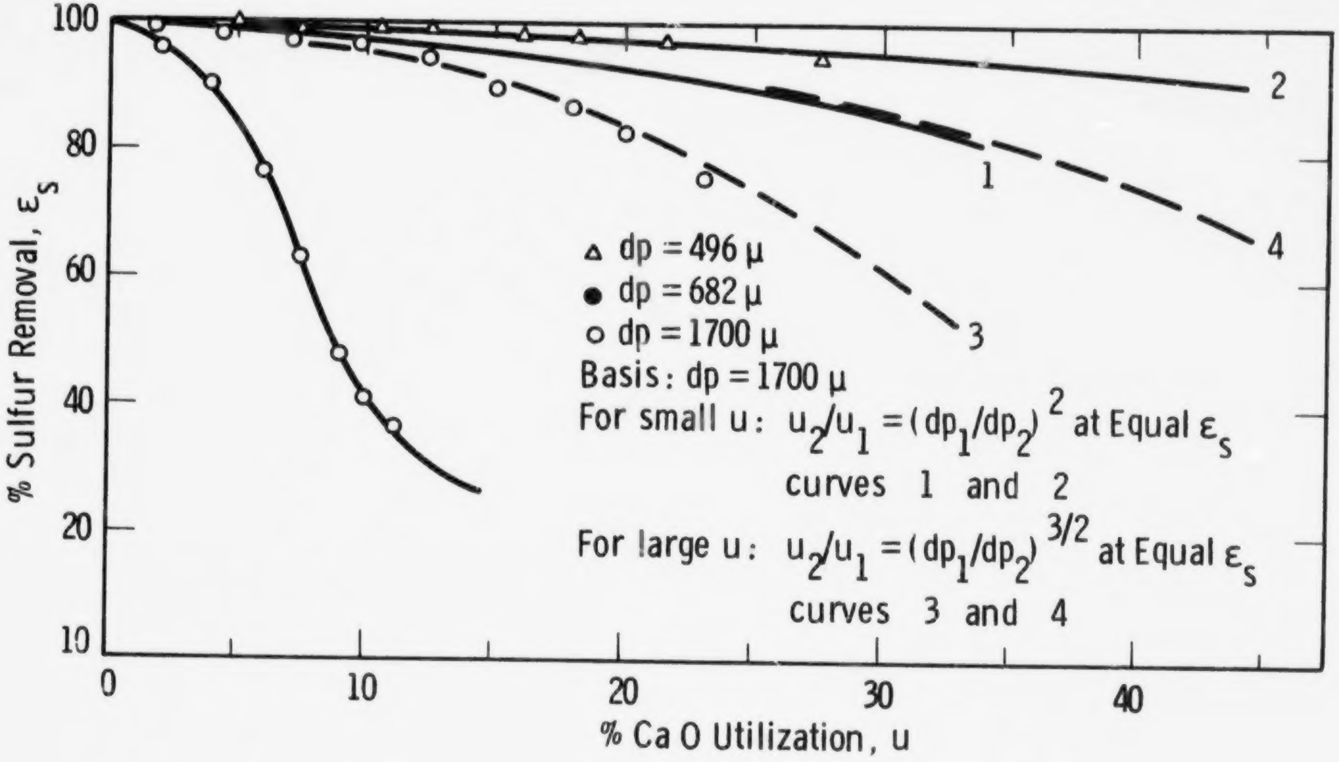

Fig. 2-4— Effect of particle size on sulfur removal

temperatures (∿ 860°C), and air/fuel ratios (∿ 24%). Each run represents a different average particle diameter, d_p, and shows the sulfur removal efficiency, ε_s, plotted against the CaO utilization, μ. The finer lime particles give greater sulfur removal for a given CaO utilization because of easier accessibility to the finer particle surfaces. The figure indicates that for small utilizations -- small enough that external particle surface is dominant in the reaction -- the utilization is inversely proportional to the average particle diameter. Or, comparing the performance of two particle sizes, 1 and 2,

$$U_2/U_1 = (d_{p_1}/d_{p_2})^2 \text{ at equal } \varepsilon_s. \tag{2-4}$$

For larger lime utilizations, where the internal particle surface enters the reaction, the proportionality is observed to be the diameter to the -3/2 power, or

$$U_2/U_1 = (d_{p_1}/d_{p_2})^{3/2} \text{ at equal } \varepsilon_s. \tag{2-5}$$

Equations (2-4) and (2-5) are demonstrated in Figure 2-4, with the experimental curve for $d_p = 1700\ \mu$ taken as the basis of comparison. Note that the point of transition between equations (2-4) and (2-5) is also a function of the average particle diameter.

The behavior of fresh bed tests in the temperature range of 830°C to 880°C may be summarized in terms of the operating variables τ_G, the gas residence time, and S, the lb CaO per lb of sulfur fed/hr. This is shown in Figure 2-5 with the sulfur removal efficiency, ε_s, plotted against the product $S\ \tau_G$ for average particle diameters 496, 682, and 1700 μ and a calcium oxide utilization equivalent to 5 wt % sulfur on the bed lime. Curves for 4 and 6 wt % sulfur on the bed lime are included for the case of $d_p = 1700\ \mu$. The behavior of Figure 2-5 is similar to that of Figure 2-4. The curves for diameters of 496 μ and 682 μ and 5 wt % sulfur on the stone are in the region where external surface dominates in the reaction, and the data appear to be consistent with the relationship

Curve 643910-A

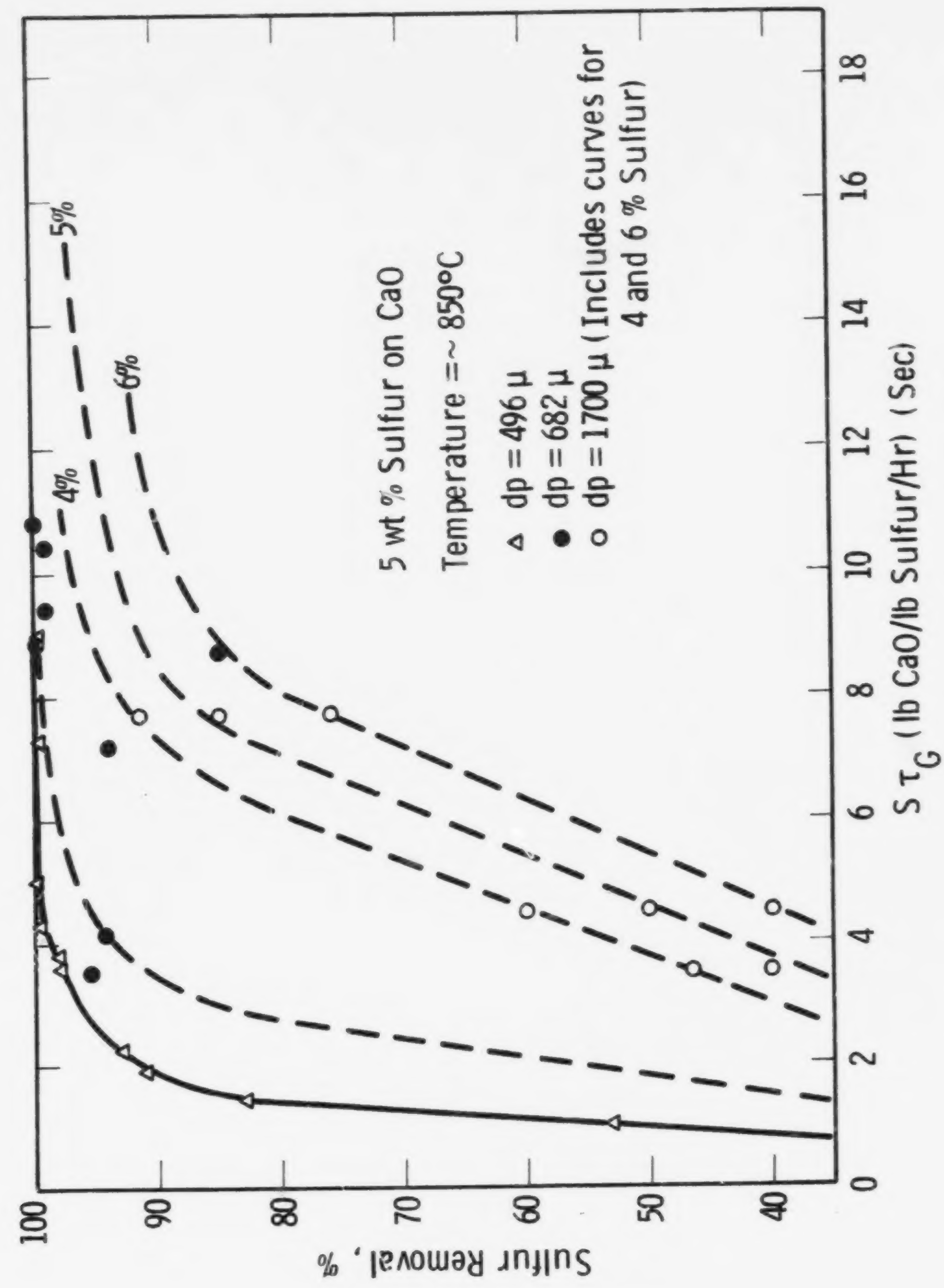

Fig. 2-5—Sulfur removal correlation

$$\frac{(S\ \tau_G)_1}{(S\ \tau_G)_2} = \left(\frac{d_{p_1}}{d_{p_2}}\right)^2 \quad \text{at equal sulfur removal.} \qquad (2\text{-}6)$$

The curve identified with the particles of 1700 μ average diameter is in the region of internal diffusion, as is seen in Figure 2-4. It may be shown that, by looking at the 682 μ particles at higher utilizations,

$$S\ \tau_G \ \alpha \ d_p^{3/2} \quad \text{at a given sulfur removal,} \qquad (2\text{-}7)$$

in the region of internal surface reaction. Figure 2-5 permits the prediction of the fresh bed operating conditions, bed depth, and velocity necessary for a given sulfur removal for a given particle size and lime utilization.

Cyclic tests differ from fresh bed tests in that the limestone make-up rate and the conditions of the stone regeneration cycle area of major importance to the sulfur removal efficiency. Except for these two factors, the behavior of cycled beds should be similar to the behavior found in fresh bed tests and summarized in Figures 2-4 and 2-5. Results of three cyclic tests are shown in Figure 2-6 as the sulfur removal versus the cycle number. The experimental conditions of the tests and the duration of the gasification and regeneration cycles are indicated on the figure. Esso (England) has observed in similar tests that the sulfur removal efficiency reaches a steady-state value of nearly 100% at a make-up rate of 1 mole of CaO per mole of sulfur fed with the U.K. limestone and the U.S. limestone 1691.

The batch, nonsteady, tests carried out by Esso (England) do not apply directly to the steady operation of a commercial size continuous unit. Fortunately, the tests indicate that extremely fast reactions are involved in the desulfurization phenomena, so that scale-up to cases with much different fluidization characteristics should be possible. For the design of a once-through operation, the fresh bed test results may be applied directly to the design, with Figure 2-5 providing a great deal of design information. The regenerative operation

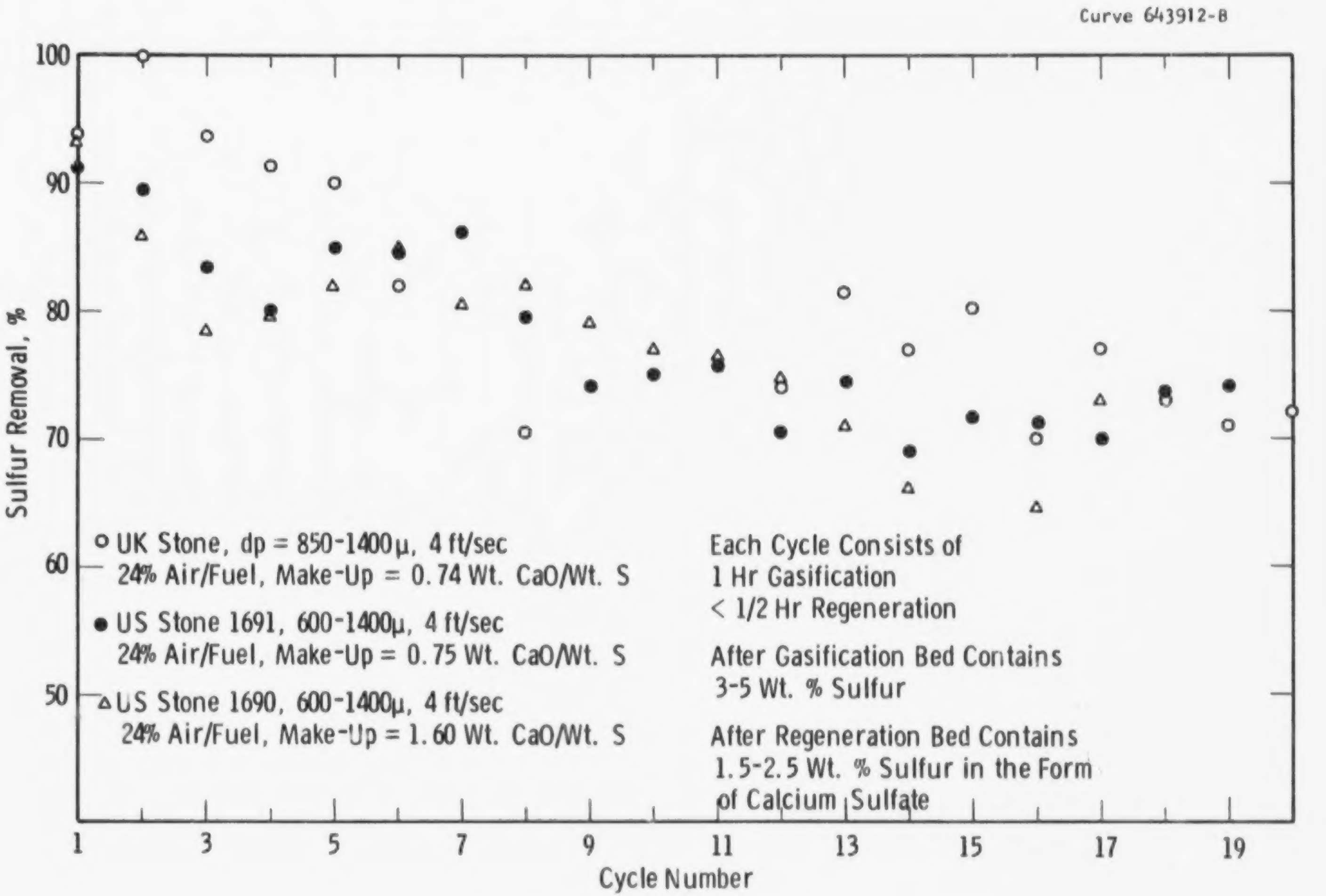

Fig. 2-6—Batch Cyclic Bed Tests

design may be somewhat less directly related to the cyclic bed tests, but the make-up rates necessary for a given sulfur removal should be valid for the continuous regenerative case, even though the regenerator performance in the batch and continuous units may differ somewhat.

Temperature levels found optimum for desulfurization in batch tests should apply to continuous operation, but the effects of the air/fuel ratio on sulfur removal observed in batch tests may differ greatly in a continuous unit in which temperature control is maintained. In addition, the gasifier temperature control scheme used in a continuous unit may itself reduce or promote sulfur removal. For example, steam or water injection in the gasifier may greatly affect the kinetics of the reaction in equation (2-3) in such a way as to reduce sulfur removal, but carbon deposition may be simultaneously reduced by the reaction

$$C + H_2O \rightleftarrows CO + H_2 \qquad (2\text{-}8)$$

to increase the sulfur removal.

Factors such as these call for further experimental investigation, but sufficient information is available from the Esso desulfurization batch data to permit preliminary design.

Regeneration

The regenerator is operated to give a slight oxygen breakthrough in the SO_2-rich product stream, thus assuring predominantly an oxidizing atmosphere in the bed. Besides converting sulfided lime back to CaO, a portion of the calcium sulfide is converted into calcium sulfate, and the carbon deposited on the lime is combusted. Thus, the dominant reactions are

$$CaS + 3/2\ O_2 \rightleftarrows CaO + SO_2 \qquad (2\text{-}9)$$

$$CaS + 2\ O_2 \rightleftarrows CaSO_4 \qquad (2\text{-}10)$$

$$C + O_2 \rightleftarrows CO_2. \qquad (2\text{-}11)$$

The main factors influencing the observed loss in lime activity may be taking place in the regeneration step. These factors are probably sintering of the stone (loss of porosity), due to the high generator temperatures, and sulfate build-up in the stone.

The mechanisms involved in batch regeneration differ somewhat from those expected in a continuous regeneration operation mainly because of differences in the temperature history of the stone in the two cases. In the batch case the whole bed heats up relatively slowly, with different reactions prevalent at different stages in the regeneration. The continuous, steady-state operation will lead to a quick heat-up of the sulfided lime particles entering the hot regenerator.

From the standpoint of regeneration of the sulfided lime and the maintenance of stone activity, about 1050°C has been found to be a suitable regenerator temperature. SO_2 concentrations of about 10 mole % have been obtained from batch regeneration runs. The SO_2 levels expected in a continuous operation will depend strongly on the rate of carbon deposition and the amount of sulfate production in the regenerator. With sulfided lime containing 3% to 5 wt % sulfur, Esso (England) has normally produced a final regenerated lime containing 1 to 2 wt % sulfur in the form of calcium sulfate. Lime residence times in the regenerator have varied from five to fifteen minutes. Esso (England) has recorded CO, CO_2, SO_2, O_2 concentrations from the regenerator and the regenerator temperature as a function of time.

The performance of a continuous regenerator is expected to be superior to that found in batch tests in terms of the amount of sulfate production in the regenerator. The operation of the Esso (England) continuous unit will supply additional information on regeneration which will be needed for more detailed design work.

Particle Carry-Over

Bed losses during the Esso (England) batch tests were caused by attrition of the particles in the calcination, gasification, and regeneration operations. The factors controlling bed losses are:

- Superficial velocity
- Quality of fluidization
- Limestone physical characteristics
- Air distributor design
- Gasifier vessel design - bed depth, freeboard height, internals, etc.
- Bed operation - recirculation of fines, turn-down.

Lime particle size distributions were measured as a function of time during the calcination, gasification, and regeneration stages of operation. Average rates of bed losses varied from 1 to 5 wt % of the bed/hr for U.K. and U.S. 1691 limestones at velocities of from 3 to 8 ft/sec, though the rate of elutriation varied greatly with time.

Since the factors listed above probably differ between the Esso (England) batch operations and a commercial power plant design, the bed losses recorded by Esso are difficult to relate to design considerations. Again, the operation of the Esso 750 kw pilot unit will provide useful information in this area.

Metal Retention

Batch tests have indicated that the bed lime removes nearly 100% of the fuel vanadium and about 20% of the fuel sodium during gasification. Tests have also indicated that this build-up of vanadium and sodium on the lime has no effect on sulfur removal, though this is not conclusive. This high metal retention behavior of the atmospheric pressure gasification/desulfurization operation should also be observed in continuous operation and is an important factor in material corrosion.

Burner Design

Esso (England) has constructed and operated a burner utilizing the clean, hot, fuel gas from batch gasification/desulfurization operations. A luminous flame, free of smoke, has been obtained with proper burner operation, indicating that the low Btu gas is easily combustible.

Preliminary Results of Continuous Operation

Operation of the Esso (England) 750 kw, regenerative, gasification/desulfurization pilot unit is now in the preliminary stages leading to the completion of the 200 hr commissioning run. Preliminary indications are that the continuous unit behaves much as expected from batch experience, with about 90% sulfur removal obtained at the expected limestone make-up rate of 1 mole CaO per mole of sulfur fed. Also, apparently because of the maintenance of the gasifier temperature, the continuous unit has been operated at much lower air/fuel ratios than were possible in the batch studies (presently down to 17% of stoichiometric). Overall behavior of the continuous unit is promising.

MATERIAL AND ENERGY BALANCES

Batch information provided by Esso (England) is taken as a basis for the preliminary design of the atmospheric pressure gasification/desulfurization concept. Both regenerative and once-through modes of operation are examined, and the temperature control schemes proposed for the systems are of prime concern. Overall material balances are computed, and the power requirements for the two modes are estimated. The preliminary design is limited to a specific set of operating conditions. Further data, however, will permit a more complete examination of the relative merits of the two modes of operation and will make possible the determination of optimum operating conditions.

Modes of Operation

Flow diagrams for both regenerative and once-through modes of operation were shown in Figure 2-1. In the regenerative operation, fuel, air, and fresh make-up limestone are introduced into the gasifier vessel to produce a hot, low-sulfur, fuel gas. The gasifier product is sent directly to the boiler, or possibly to cyclone clean-up before being transported to the boiler. The hot sulfided lime produced in the gasifier is sent to the regenerator, converted to a less active form of lime, and returned to the gasifier. The SO_2 released from the regenerator is transported to a sulfur recovery system, possibly yielding a useful by-product to reduce the system operating cost. In the regenerative system, the control of the gasifier and regenerator temperatures is interdependent with the regenerator temperature control scheme determining the temperature control requirements in the gasifier; and the carbon deposition rate is an important factor in the overall operation of the system.

The once-through mode of operation differs from the regenerative operation in that the lime is utilized to a greater extent in the gasifier and is disposed of after being converted to the sulfate form in the sulfate generator. Temperature control of the gasifier and sulfate generator is an important consideration, as in the regenerative mode, but the carbon deposition rate is less important to the overall once-through system operation.

Design Parameters

Design parameters of importance in oil gasification/desulfurization are listed in the following matrix. The four factors listed as process specifications (geographic location, fuel oil, limestone, and sulfur removal) are determined by the market outlook, by residual oil and limestone availability, and by the expected SO_2 emission standards, respectively.

The design variables listed refer to the broad design factors which will eventually determine the feasibility of preliminary design concepts. The design considered may be that of a new boiler or a boiler retrofit, while three factors -- boiler size, boiler load factor, and boiler turn-down ratio -- are important in determining the design for either a new system or a retrofit system. In the retrofit case the type of boiler being retrofit is important because coal- and oil-fired boilers differ in retrofit characteristics. The location of the gasification system relative to the boiler structure (internal or external to the boiler) and the configuration of the gasification system (singular or modular) are basic considerations in both new and retrofit designs. The other important design variable considered in the preliminary design is the mode of operation (regenerative or once-through).

The operating variables listed are those found to be important in the Esso (England) batch studies. In addition, the schemes proposed to control the temperatures of the gasifier, regenerator, and sulfate generator are listed. Possible methods of controlling gasifier temperature are the recycle of stack gas (presently being used with the Esso continuous pilot unit), water or steam injection, the use of heat transfer surface in the gasifier to preheat boiler water or, possibly, reducing the air/fuel ratio to the point of thermal balance. In the regenerative operation, the regenerator temperature might be controlled

DESIGN PARAMETERS

PROCESS SPECIFICATIONS
- Geographic location
- Fuel oil
- Limestone
- Sulfur removal

DESIGN VARIABLES

	Boiler Size	Load Factor	Turn-Down Ratio
New boiler design Retrofit design - (coal- or oil-fired)			

- Gasification system location
  - Internal to boiler
  - External to boiler

- Gasification system operational mode
  - Regenerative or once-through

- Gasification system configuration
  - Unit design or modular, etc.

OPERATING VARIABLES
- Fluidization velocity
- Particle sizes
- Air/Fuel ratio
- Limestone make-up rate
- Limestone utilization
- Bed depth
- Bed temperature - gasifier and regenerator

TEMPERATURE CONTROL
- Gasifier temperature control
  - Stack gas recycle
  - Water injection
  - Steam injection
  - Heat transfer surface
  - Air/fuel ratio

- Regenerator temperature control
  - Stone circulation rate
  - Addition of make-up stone
  - Heat transfer surface

- Sulfate generator temperature control
  - Excess air circulation
  - Heat transfer surface

by regulating the lime circulation rate (as in the Esso continuous pilot unit), by adding fresh make-up limestone to the regenerator, or by including heat transfer surface in the regenerator vessel. Methods proposed for control of the sulfate generator in the once-through operation are the circulation of excess air through the sulfate generator or the inclusion of heat transfer surface in the vessel.

The specifications assumed for the conceptual design are presented in the next matrix. The process specifications include a geographic location for the commercial development of the atmospheric pressure residual oil gasification/desulfurization process in the Eastern United States according to market predictions. Preliminary designs are based on a residual oil containing 3 wt % sulfur with low heating value of 17,700 Btu/lb, a limestone having the composition of the U.K. limestone investigated by Esso (England) (Table 2-2), and a sulfur removal efficiency of 90% to 95% to meet present and future pollution standards.

For cost estimates, a boiler capacity of 600 MW and load factors of 40% and 80% were considered. In general, capacities of 100 MW and greater are of interest in the preliminary designs. A turn-down ratio of 4/1 was of interest in both new and retrofit designs.

The operating variables are specified according to the Esso (England) batch data. These data indicate that the gasifier temperature of 1600°F is optimum for both regenerative and once-through operations from the standpoint of sulfur removal. From the standpoint of lime regeneration, the data indicate 1900°F for the regenerator temperature. A temperature of 1500°F is chosen for the sulfate generator, based on chemical equilibrium to give an SO_2 partial pressure of less than 0.01 atm coming off the generator, although the kinetics of the reaction have not been investigated. Based on the Esso (England) cyclic batch tests a limestone make-up rate of 1 mole CaO per mole of sulfur fed is specified for 90% to 95% sulfur removal in the regenerative operation, with 5 wt % sulfur on the gasifier lime, an average stone diameter of 2000 μ, a

PROCESS SPECIFICATIONS

Geographic location: Eastern United States
Fuel oil: 3 wt. % sulfur; LHV = 17,700 Btu/lb.
Limestone: U.K. limestone
Sulfur removal: 90-95%

DESIGN VARIABLES

Boiler size: 600 MW
Load factor: 40%, 80%
Turn-down: 4/1

OPERATING VARIABLES

	REGENERATIVE	ONCE-THROUGH
Gasifier temperature	1600°F	1600°F
Regenerator (sulfate generator) temperature	1900°F	1500°F
Stone make-up rate	1 mole CaO/mole sulfur	3 moles CaO/mole sulfur
Air/fuel ratio	20% of stoichiometric	20% of stoichiometric
Limestone utilization	∿ 5 wt % sulfur in bed	∿ 19 wt % sulfur in bed
Fluidization velocity	8 ft /sec	8 ft /sec
Particle sizes (avg.)	∿ 2000 μ	∿ 1000 μ
Gasifier bed depth (static)	2.5-3.5 ft	3.5-4.0 ft

PROJECTED PERFORMANCE

	REGENERATIVE	ONCE-THROUGH
Sulfur removal	90-95%	90-95%
Thermal efficiency	90-98	90-98
Elutriation	∿ 2-5% of bed/hr	∿ 2-5% of bed/hr
Vanadium retention	∿ 100%	∿ 100%
Sodium retention	∿ 20%	∿ 20%

velocity of 8 ft/sec in the gasifier, and a gasifier bed depth of 3.5 to 4.0 ft. An air/fuel ratio of 20% of stoichiometric is specified from the Esso data in order to assure reasonable carbon deposition rates in the regenerative and once-through operations.

The projected performance for regenerative and once-through operations is identical and is based on the batch experimental results provided by Esso (England).

Temperature Control

Energy and material balances are carried out to determine the requirements and feasibility of the methods proposed for controlling the temperatures of the regenerative and once-through systems. Factors such as the system thermal efficiency, gasifier and regenerator product characteristics, apparatus dimensions, and temperature control requirements are investigated with respect to the proposed temperature control methods as a function of the carbon deposition rate. The advantages and disadvantages of the proposed temperature control schemes are compared.

Regenerative Operation

Control of the gasifier temperature at 1600°F and the regenerator temperature at 1900°F is interdependent, with the temperature control scheme used in the regenerator affecting the temperature control requirements in the gasifier. The performance of the regenerative operation is highly dependent on the temperature control schemes applied to the system and the rate of carbon deposition occurring in the gasifier.

Regenerator

The methods proposed for controlling the regenerator temperature at 1900°F are regulation of the rate of lime circulation between the gasifier and regenerator, addition of fresh make-up limestone to the regenerator, or inclusion of heat transfer surface in the regenerator vessel to preheat boiler water.

Heat is liberated in the regenerator by the three reactions

$$CaS + 3/2\ O_2 \rightleftarrows CaO + SO_2,\ \Delta H_1 \qquad (2\text{-}9)$$

$$CaS + 2\ O_2 \rightleftarrows CaSO_4,\ \Delta H_2 \qquad (2\text{-}10)$$

$$C + O_2 \rightleftarrows CO_2,\ \Delta H_3, \qquad (2\text{-}11)$$

where ΔH_1, ΔH_2, and ΔH_3 are the heats of reactions of equations (2-9), (2-10), and (2-11), respectively, in Btu/lb-mole. The following assumptions are made in evaluating the regenerator temperature control requirements:

- Per pound of fuel entering the gasifier, $Y\varepsilon_s$ lbs of sulfur enter the regenerator (where Y is the wt fraction of sulfur in the fuel and ε_s is the gasifier sulfur removal efficiency), and 90% of this entering sulfur is converted to SO_2 by reaction (2-9) while the remaining 10% is converted to calcium sulfate by reaction (2-10).
- Deposited carbon enters the regenerator at a rate of $X_c N_G$ (lb carbon/lb fuel) where X_c is the weight fraction of carbon of the gasifier lime and N_G is the rate of lime circulation to the regenerator (lb lime/hr per lb of fuel/hr). All of the deposited carbon is combusted by reaction (2-11).
- Fresh make-up limestone is added to the regenerator at a rate of N_{MR} (lb limestone/hr per lb fuel/hr) and is calcined completely by the reaction

  $$CaCO_3 \rightleftarrows CaO + CO_2 \ , \ \Delta H_4 \qquad (2\text{-}12)$$

  having a heat of reaction ΔH_4.
- Sulfided lime enters the regenerator with a temperature of 1600°F and is withdrawn at a temperature of 1900°F, the lime having a heat capacity, C_{p_s}, of 0.3 Btu/lb °F. Air and fresh limestone enter the regenerator at 77°F and exit at 1900°F. A stoichiometric amount of air is fed to the regenerator.
- Heat losses to the surroundings are neglected.

An energy balance written around the regenerator indicates that for adiabatic operation

$$0 = \Delta H_1\ (.9)\ \frac{(Y)(\varepsilon_s)}{32} + \Delta H_2\ (.1)\ \frac{(Y)(\varepsilon_s)}{32} + \Delta H_3\ \frac{X_c N_G}{12}$$

$$+ \Delta H_4\ (.434)\ \frac{N_{MR}}{44} + N_G\ C_{P_s}\ (300) + N_{MR}\ C_{P_s}\ (.566)\ (1823)$$

$$+ F_{N_2}\ C_{P_{N_2}}\ (1823) + F_{CO_2}\ C_{P_{CO_2}}\ (1823) + F_{SO_2}\ C_{P_{SO_2}}\ (1823) \qquad (2\text{-}13)$$

where F_{N_2}, F_{CO_2}, and F_{SO_2} are the molar flow rates of nitrogen, carbon dioxide, and sulfur dioxide per lb of fuel, respectively, and Cp_{N_2}, Cp_{CO_2}, and Cp_{SO_2} are the heat capacities of N_2, CO_2, and SO_2, respectively. F_{N_2}, F_{CO_2}, and F_{SO_2} are given by

$$F_{N_2} = \frac{X_c N_G}{12} + 1.5\ (.9)\ \frac{Y\varepsilon_s}{32} + 0.1\ \frac{Y\varepsilon_s}{32}\ , \qquad (2\text{-}14)$$

$$F_{CO_2} = \frac{.434\ N_{MR}}{44} + \frac{X_c N_G}{12}\ , \qquad (2\text{-}15)$$

$$F_{SO_2} = (.9)\ \frac{Y\varepsilon_s}{32}\ . \qquad (2\text{-}16)$$

Placing the proper values in equation (2-13) and solving for N_G, the lime circulation rate per lb of fuel/hr, gives

$$N_G = 88\ X_c N_G + 41.3\ Y\varepsilon_s - 14.5\ N_{MR}. \qquad (2\text{-}17)$$

Equation (2-17) relates the flow rate of sulfided lime necessary for temperature control of the regenerator to the carbon deposition rate, $X_c N_G$, the rate of sulfur removal per lb of fuel/hr, $Y\varepsilon_s$, and the rate of addition of make-up stone to the regenerator per lb of fuel/hr, N_{MR}.

On the basis of 95% sulfur removal, or $\varepsilon_s = 0.95$, equation (2-17) is illustrated in Figure 2-7. Equation (2-17) becomes

$$N_G = 88\ X_c N_G + 39.3\ (Y - 0.369\ N_{MR}) \tag{2-18}$$

and Figure 2-7 is a plot of N_G versus $X_c N_G$ with $Y-0.369\ N_{MR}$ as a parameter. Curves of constant X_c are also shown on Figure 2-7 (as broken lines) to reflect the behavior expressed by equation (2-2). That is, X_c is a constant independent of N_G for $N_G \ll K_2$. Two schemes of regenerator temperature control are incorporated in the figure: control by sulfided lime circulation (represented by $N_{MR} = 0$), and control by addition of fresh make-up limestone to the regenerator (represented by $N_G = N_{G_{MIN}}$, the minimum circulation rate necessary to maintain the maximum allowable gasifier lime utilization). The minimum circulation rate is given by

$$N_{G_{MIN}} = \frac{Y\varepsilon_s}{X_{SG} - X_{SR}} \tag{2-19}$$

where X_{SG} is the maximum weight fraction of sulfur on the gasifier lime in order to achieve a sulfur removal of ε_s, and X_{SR} is the weight fraction of sulfur on the regenerator lime, in the form of calcium sulfate. For $Y = 0.03$, $\varepsilon_s = 0.95$, X_{SG} must be 0.05 according to the Esso (England) data, and X_{SR} is assumed to be about 0.005, giving a $N_{G_{MIN}}$ of 0.633 lb lime/lb fuel.

Figure 2-7 indicates that the control value of N_G can easily rise to three or four times the minimum value, $N_{G_{MIN}}$, while relatively small amounts of fresh make-up limestone can be used to accomplish the temperature control ($N_{MR} = 0.05$ lb/lb fuel with $X_c = 0.003$, $Y = 0.03$, and $N_G = N_{G_{MIN}}$). In fact, equation (2-17) indicates that the critical value of X_c above which temperature control of the regenerator can no longer be accomplished with $N_{MR} = 0$ is 0.0114. That is, for $X_c > 0.0114$, an increase in N_G will have no effect on the net rate at which heat is generated in the regenerator. This conclusion holds only if the air is

Curve 643987-B

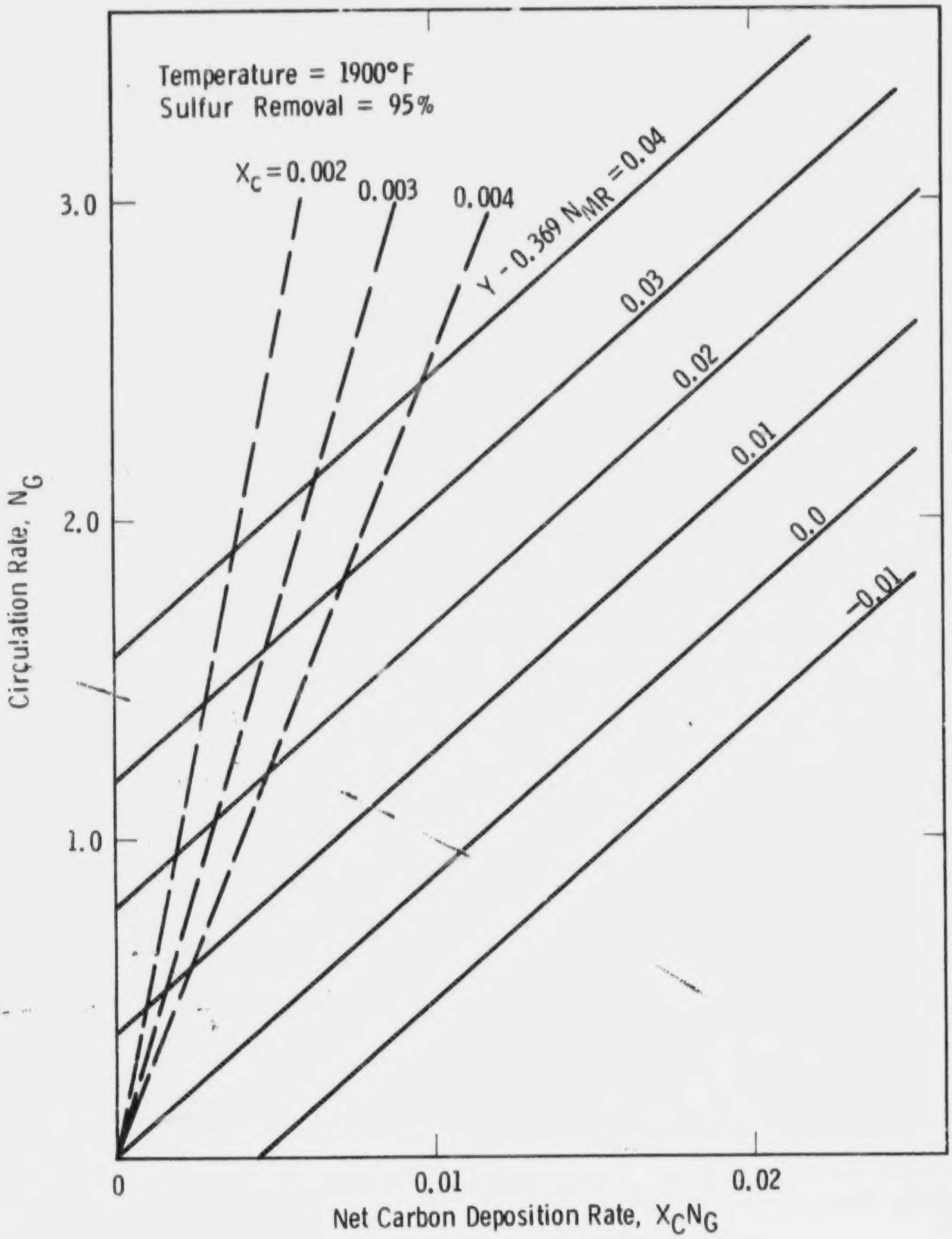

Fig. 2-7—Regenerator Temperature Control

kept at the stoichiometric feed rate to give complete combustion of the deposited carbon at all values of X_c. It is necessary to operate in this manner, though, to ensure regeneration of the lime.

The third temperature control scheme, the use of heat transfer surface in the regenerator with $N_G = N_{G_{MIN}}$, is limited by the wall-surface to volume ratio in the regenerator. Based on the air rate to the regenerator at 1900°F, a superficial velocity of 8 ft/sec, 95% sulfur removal, a gasifier thermal efficiency of 95%, and a plant heat rate of 10^4 Btu/kwh, the cross-sectional area of the regenerator per MW is shown in Figure 2-8 versus the carbon deposition rate, with the weight % of sulfur in the fuel as a parameter. The regenerator cross-sectional area is strongly dependent on the carbon deposition rate, and assuming an expanded bed height of about three feet, the possibility of using a water-jacketed regenerator for temperature control appears limited to small units. A 20 MW unit would require a reasonable heat transfer coefficient of about 25 Btu/hr-ft^2-°F for temperature control based on the wall-surface area of the vessel, while a 200 MW unit would require a heat transfer coefficient of about 90 Btu/hr-ft^2-°F. The additional surface available with a water-cooled air distributor in the regenerator does not appear sufficient to give reasonable heat transfer coefficients with capacities greater than 200 MW. Thus, the inclusion of heat transfer surface internal to the fluid bed, or the restriction to modular regenerator designs, must be considered in order to use this temperature control scheme.

Another important factor in regenerator temperature control is the concentration of SO_2 leaving the regenerator. As the carbon deposition rate, X_cN_G, increases, the molar concentration of SO_2 leaving the regenerator will decrease, making sulfur recovery more expensive. From equations (2-14) through (2-16) the SO_2 concentration is given by

$$X_{SO_2} = \frac{Y\varepsilon_s}{12.7\ X_cN_G + 6.64\ Y\varepsilon_s + 0.316\ N_{MR}}\ , \qquad (2\text{-}20)$$

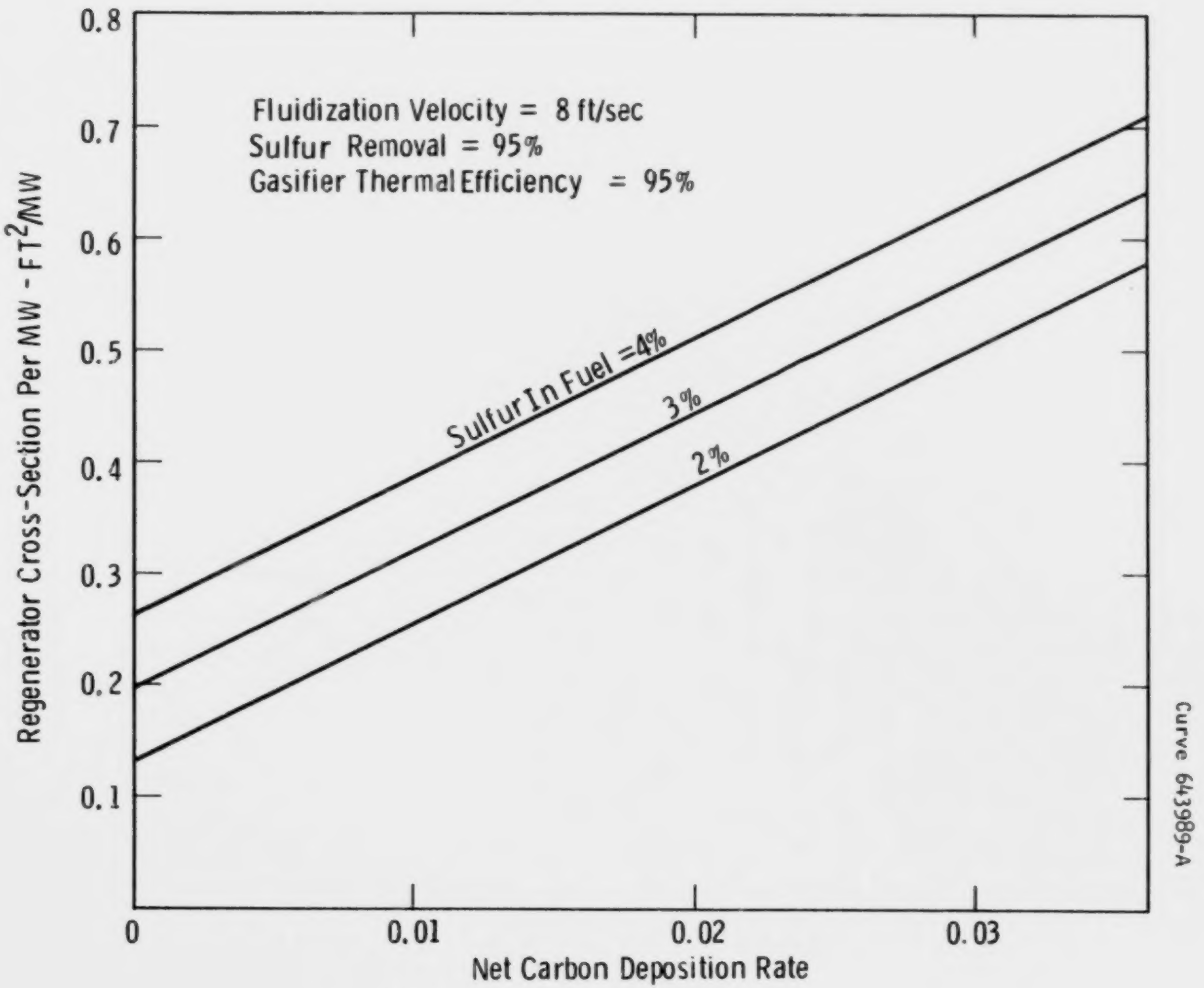

Fig. 2-8—Regenerator Design

where X_{SO_2} is the mole fraction of SO_2 in the regenerator product. Equation (2-20) is shown in Figure 2-9. Since the temperature control scheme "$N_{MR} = 0$" operates with a high carbon deposition rate, the SO_2 concentration will be reduced below that of the other two schemes. For instance, if $X_c = 0.004$ and $Y = 0.03$, then Figure 2-7 indicates that for temperature control with $N_{MR} = 0$ the carbon deposition rate will be almost 0.01 lb carbon/lb fuel, while with $N_G = N_{G_{MIN}}$ the deposition rate will be 0.0025 lb carbon/lb fuel. Then from Figure 2-9 the mole fraction of SO_2 from the regenerator will be 0.090 for $N_{MR} = 0$, 0.130 for $N_G = N_{G_{MIN}}$, and 0.130 for control with heat transfer surface.

The advantages and disadvantages of the three regenerator temperature control schemes are compared in the next matrix. The points summarized there indicate that:

- The use of stone circulation for regenerator temperature control is attractive only if the amount of carbon deposited on the gasifier lime does not exceed the value of $X_c \approx 0.004$ and if the carbon deposition phenomenon is well-behaved and predictable (that is, the carbon deposition must not fluctuate during operation and must be a predictable quantity for design purposes).
- The other two temperature control schemes become more attractive than stone circulation for $X_c \overset{\sim}{>} 0.004$.
- The use of make-up limestone in the regenerator yields a high SO_2 concentration in the regenerator product and greatly reduces the stone circulation rate from the value which would be needed with the stone circulation temperature control method. If the carbon deposition phenomenon is not well-behaved, then the specified make-up rate of 1 mole CaO/mole sulfur fed must be sent totally to the gasifier to achieve the prescribed sulfur removal and minimize the control problem involved in solids flow while the make-up necessary for regenerator temperature

Curve 646232-A

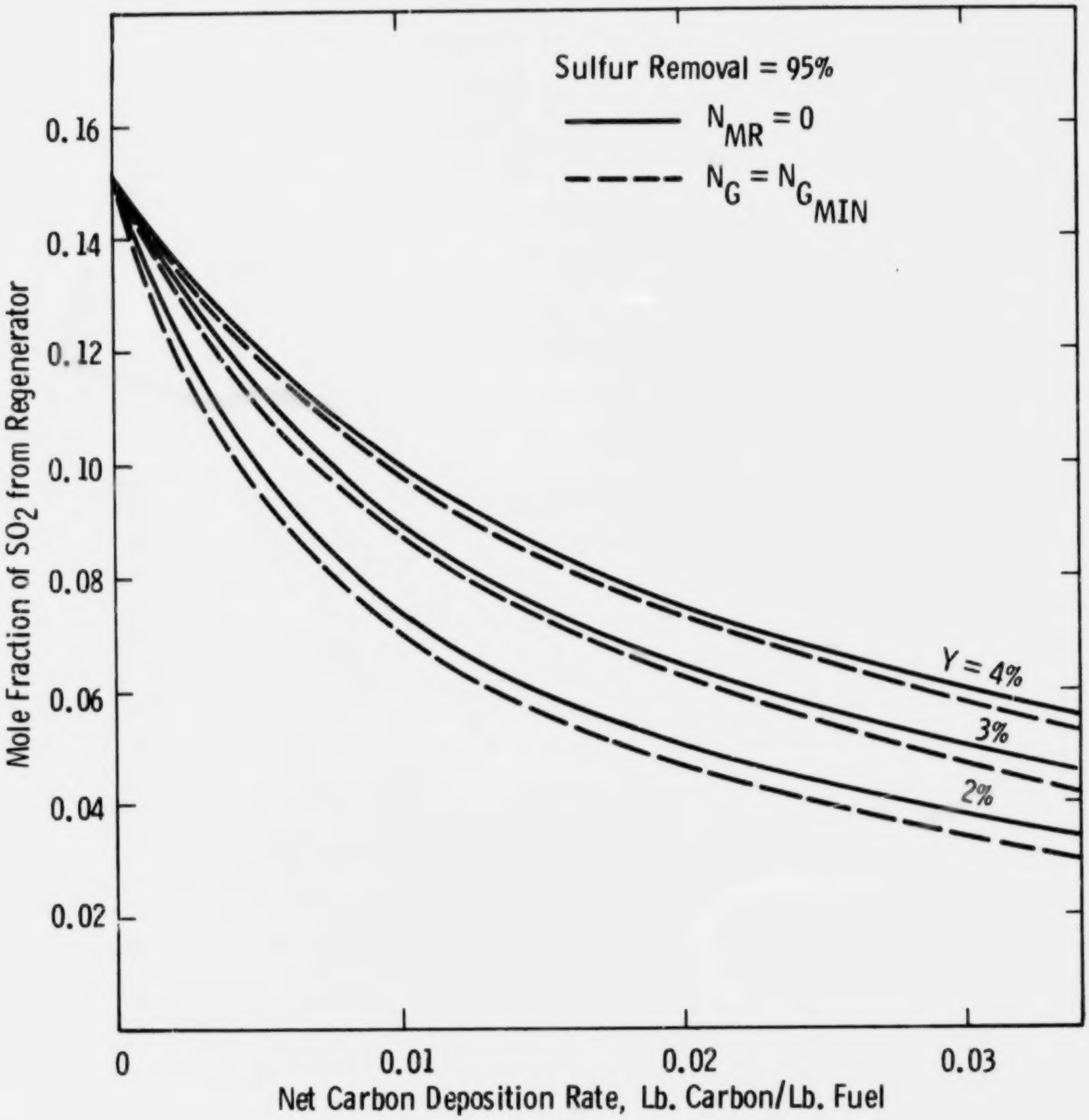

Fig. 2-9—Regenerator Operation

COMPARISON OF REGENERATOR TEMPERATURE CONTROL SCHEMES

CONTROL SCHEME	ADVANTAGES	DISADVANTAGES
Stone circulation ($N_{MR} = 0$)	Circulation of stone is already a necessary element of operation A simple scheme if the carbon deposition behavior is predictable Best of the three schemes for $X_c \tilde{<} 0.002$	May necessitate extremely large circulation rates for which capital and operating expenses are high. (At $X_c = 0.005$, $Y = 0.03$, and $\varepsilon_s = 0.95$, $N_G = 2$ lb stone/lb fuel) Gives higher carbon carry-over to the regenerator than other schemes, decreasing the SO_2 concentration from the regenerator (a probable decrease in X_{SO_2} of about 25% as compared to other schemes) Makes gasifier and regenerator performance interdependent
Make-Up stone ($N_G = N_{G_{MIN}}$)	Minimizes stone circulation and carbon deposition rate ($N_{G_{MIN}} = 0.633$ with $Y = 0.03$ and $\varepsilon_s = 0.95$) Higher SO_2 concentration than with stone circulation scheme. ($X_{SO_2} \approx 0.12$)	Calls for increase in the amount of make-up stone fed to system (make-up rate to gasifier ≈ 0.09 lb/lb fuel, while $N_{MR} \approx 0.05$ lb/lb fuel) Slight decrease in thermal efficiency compared to other schemes (~ 1% less efficient)
Heat transfer surface (H.T.S.)	Minimizes stone circulation and carbon deposition rate ($N_G = N_{G_{MIN}}$) Highest SO_2 concentration and highest thermal efficiency of the three schemes. ($X_{SO_2} \approx 0.12$, thermal efficiency $\approx 0.97\%$ depending on gasifier control scheme)	Control is complex High regenerator capital cost May be feasible only with modular regenerator design of H.T.S. internal to the bed in systems > 100 MW

control is added to the regenerator. This requires a greater consumption of limestone and, thus, a higher operating cost. If the carbon deposition is a well-behaved phenomenon then the total make-up rate of 1 mole CaO per mole of sulfur fed may be distributed between the gasifier and regenerator to maintain regenerator temperature control and remove the specified amount of fuel sulfur. Thus, the rate of limestone consumption might be minimized in this case. This indicates the need for further experimental investigation of carbon deposition during continuous operation.

- The use of heat transfer surface maximizes the SO_2 concentration relative to the other two control methods, but it may be associated with high capital costs and may be limited in feasibility.

Gasifier-Regenerative Operation

The gasifier temperature may be maintained at 1600°F by one of five methods: stack gas recycle, steam injection, water injection, heat transfer surface, or operating at the minimum air/fuel ratio. The first three depend on the addition of a heat storage component, stack gas, steam, or water, to the gasifier to carry the excess heat generated in the gasifier as sensible heat. In this way no heat is lost, but the system equipment is enlarged. A penalty must be paid in the steam and water injection cases, since the heat of vaporization of water is nonrecoverable. Also, stack gas, steam, and water are not inert to the gasification/desulfurization phenomena. In particular, steam and water injection may lead to a reduction in the rate of carbon deposition by the reaction

$$C + H_2O \rightleftarrows CO + H_2. \qquad (2\text{-}21)$$

On the other hand, the presence of large excesses of H_2O in the gasifier may lead to a reduction in sulfur removal by reaction (2-3). The sudden evaporation of injected water droplets may also contribute to the distribution of injected residual oil in the fluidized bed upon hitting the hot gasifier lime. Further experimental studies must be carried out to investigate the effects of stack gas, steam, and water on the gasification/desulfurization operation. For the purposes of preliminary design it is assumed that these temperature control components behave as inerts.

The inclusion of heat transfer surface in the gasifier will allow the system to operate with high thermal efficiency if boiler water is preheated in the gasifier, and will lead to a reduction in equipment sizes over those of the first three proposed methods. The cost of the gasifier will increase greatly, and a detailed cost analysis of the trade-off between heat transfer surface and stack gas recycle, steam injection, or water injection must be performed to evaluate the relative merits of these proposed temperature control methods.

If the gasification/desulfurization operation can be carried out at a low enough air/fuel ratio (about 14% of the stoichiometric air rate) then the gasifier may be thermally balanced at 1600°F so that further temperature control consideration is unnecessary. Operating at this air/fuel ratio would minimize equipment sizes, control problems, and overall system complexity. It has not been demonstrated that operation with the minimum air/fuel ratio can be realized, and at such a low air/fuel ratio the desulfurization may be disrupted by high carbon deposition rates. In any case, the lowest air/fuel ratio which avoids excess carbon deposition is the most reasonable choice for this operating variable. For preliminary design it is assumed that the minimum air/fuel ratio may be applied to the gasification/desulfurization process without increasing the carbon deposition rate above that expected at 20% of stoichiometric air.

In the regenerative operation, the choice of regenerator temperature control scheme affects the temperature requirements in the gasifier and the overall system performance. This is demonstrated by examining the gasifier thermal efficiency and the gasifier product hot gas heating value as a function of the carbon deposition rate and the gasifier and regenerator temperature control methods.

Energy and material balances are carried out around the gasifier in terms of the following quantities: $\Delta H_G \equiv$ the heat generation rate at 77°F in the gasifier due to cracking, combustion, and desulfurization, in Btu/lb fuel (as in Figure 2-3); $F_p \equiv$ mass flow rate of gasifier product, excluding the temperature control stream, in lb/hr/lb fuel/hr; $\Delta H_p \equiv$ the enthalpy change in the gasifier product going from 77°F to 1600°F, in Btu/lb, $= -C_p$ 1523°F (from the Esso data $C_p \simeq 0.35$ Btu/lb°F); $F_S \equiv$ mass flow rate of the temperature control stream (stack gas, steam, or water) in lb/hr/lb fuel/hr; $\Delta H_S \equiv$ the enthalpy change in the temperature control stream going from its initial conditions to 1600°F ($\Delta H_S \simeq 1800$ Btu/lb water based on water at 77°F, 710 Btu/lb steam based on saturated steam at 1 atm, and 358 Btu/lb stack gas based on stack gas at 300°F). The limestone make-up rate to the gasifier, N_{MG}, is given by

$$N_{MG} = \frac{Y}{32} \frac{56}{.552} = 3.17\ Y \text{ lb limestone/hr/lb fuel/hr} \qquad (2\text{-}22)$$

based on 1 mole of CaO per mole of sulfur fed. Then, the limestone contributions to the gasifier energy balance are

(1) Energy to calcinate the make-up limestone in the gasifier

$= 0.434\ (3.17\ Y)\ 1740 = 2397\ Y$ Btu/lb fuel.

(2) Energy to heat the make-up lime from 77°F to 1600°F

$= 0.3\ (3.17\ Y)\ 0.566\ (1523) = 820\ Y$ Btu/lb fuel.

(3) Energy to heat the CO_2 calcination product from 77°F to 1600°F

$$= 0.434\ (N_{MG})\ \frac{12.0}{44}\ (1523) = 573\ Y\ \text{Btu/lb fuel.}$$

(4) Energy returned from the regenerator to the gasifier as

1900°F lime = $0.3\ [N_G - 0.566\ N_{MG}]\ 300$ Btu/lb fuel.

(5) The energy lost through the gasifier boundaries is assumed to be 1% of the fuel oil heating value per lb of fuel fed

$$= 177\ \text{Btu/lb fuel.}$$

The combustion value of material taken from the gasifier to the regenerator is given by

(6) Sulfur lost = $\epsilon_S Y\ (10{,}630)$ Btu/lb fuel;

(7) Carbon lost = $X_c N_G\ (14{,}093)$ Btu/lb fuel.

An energy balance around the gasifier results in

$$F_S \Delta H_S = \Delta H_G + 90\ \{N_G - 0.566\ N_{MG}\} + F_P \Delta H_P - 3790\ Y - 177 \quad (2\text{-}23)$$

where

$$3790\ Y \equiv [1] + [2] + [3].$$

The temperature control requirements, F_S, for stack gas recycle, steam injection, or water injection can be calculated from equation (2-23) for a given air/fuel ratio, sulfur content in the residual oil fuel, sulfur removal efficiency, carbon deposition rate, regenerator temperature control method, and gasifier control method.

The total loss in gasifier product heating value is given by the sum of items (1), (2), - (4), (5), (6), (7) and $F_S\ \lambda$, where the last item applies to the case of steam injection and water injection with λ being the heat of vaporization of water = 980 Btu/lb. Then for all five proposed gasifier temperature control schemes the thermal efficiency, ε_T, is given by

$$1 - \varepsilon_T = \frac{(1) + (2) - (4) + (5) + (6) + (7) + F_S\lambda}{17,700}$$

$$1 - \varepsilon_T = [\varepsilon_S Y\ (10,630) + X_c N_G\ (14,093) - 90\ [N_G - 0.566\ N_{MG}]$$

$$+ 3217\ Y + 177 + F_S\lambda]/17,700 \qquad (2\text{-}24)$$

with regenerator temperature control by lime circulation or by addition of fresh stone to the regenerator. In the case of heat transfer surface in the regenerator, the thermal efficiency is identical to that of the lime circulation method. Equation (2-24) may be used to evaluate the thermal efficiency for a given air/fuel ratio, sulfur content in the residual oil fuel, sulfur removal efficiency, carbon deposition rate, regenerator temperature control method, and gasifier temperature control method.

Figures 2-10 and 2-11 illustrate equation (2-24) as a function of the carbon deposition rate for air/fuel ratios of 20% and 25% of stoichiometric with 95% sulfur removal of a 3 wt % sulfur fuel oil. Curves for 2 and 4 wt % sulfur fuel oils are very similar to Figures 2-10 and 2-11. Note that the thermal efficiency is the same for the three gasifier temperature control schemes in which a heat of vaporization penalty is not paid: stack gas recycle, heat transfer surface, and minimum air/fuel ratios (∿ 14% of the stoichiometric air rate). The thermal efficiency is also independent of the air/fuel ratio for stack gas recycle and heat transfer surface temperature control schemes, while increasing the air/fuel ratio lowers the thermal efficiency in the steam and water injection temperature control methods. Increasing the carbon deposition rate lowers the thermal efficiency with all five of the proposed temperature control schemes unless the cases of steam and water injection act to remove deposited carbon in the gasifier by

Curve 643985-B

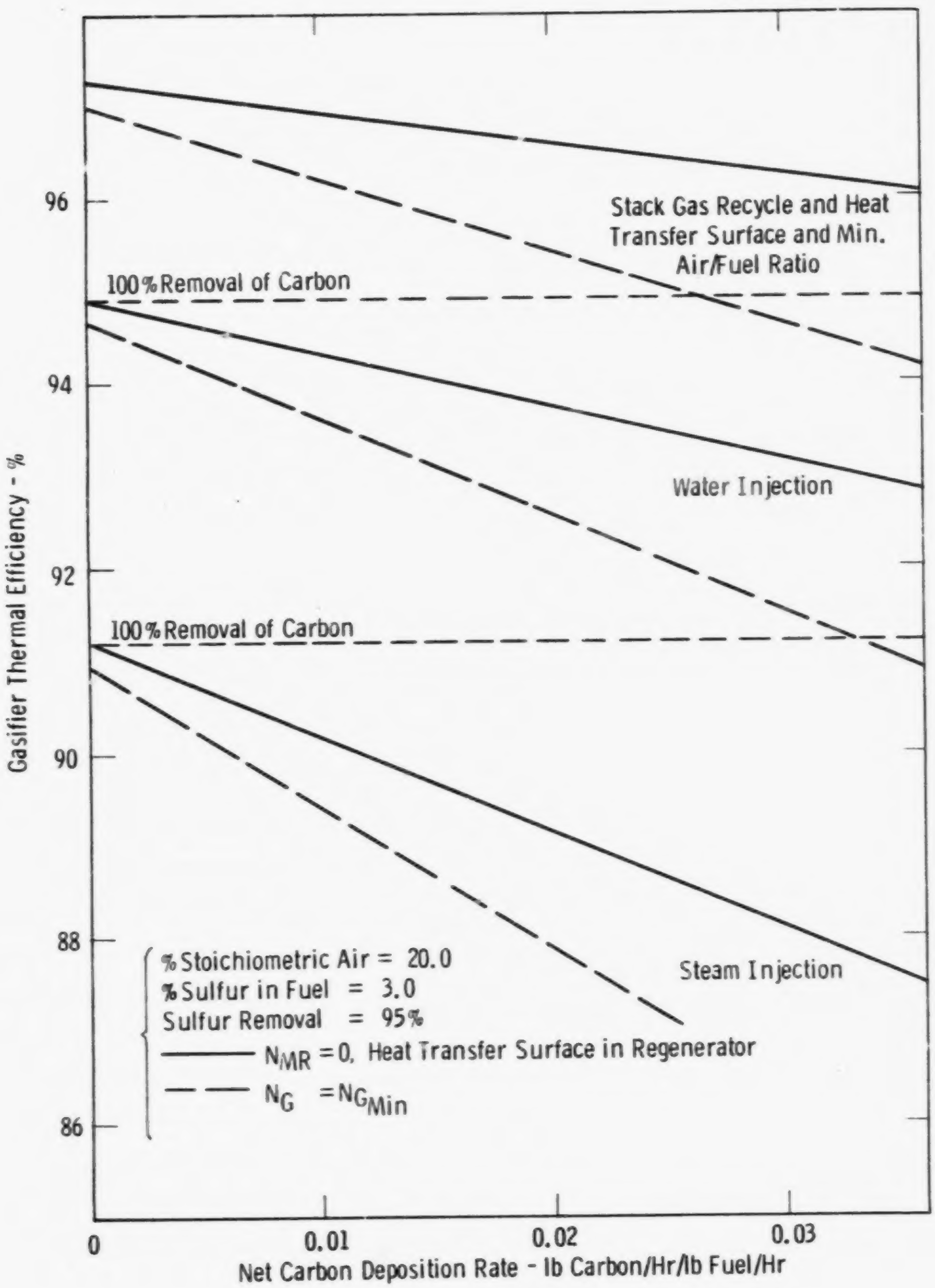

Fig. 2-10—Gasifier Temperature Control

Curve 643986-B

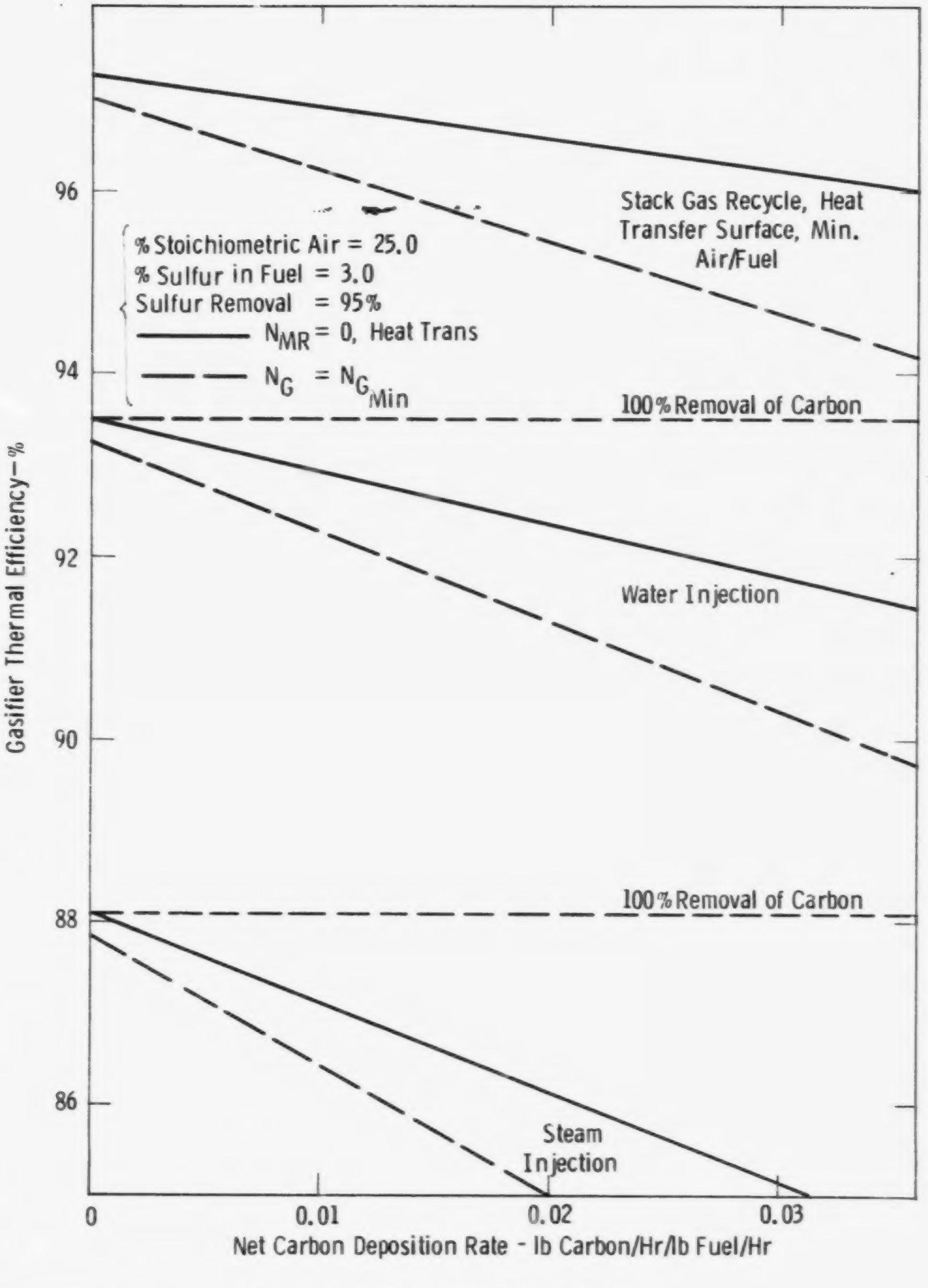

Fig. 2-11—Gasifier Temperature Control

reaction (2-21). The situation in which 100% of the deposited carbon is removed by reaction (2-21) is indicated in Figures 2-10 and 2-11 by the broken lines labeled "100% removal of carbon."

The method of regenerator temperature control also affects the gasifier thermal efficiency, with the method of feeding make-up limestone to the regenerator ($N_G = N_{G_{MIN}}$) giving lower thermal efficiencies than the stone circulation ($N_{MR} = 0$) or heat transfer surface methods for a given carbon deposition rate. The methods using make-up limestone and heat transfer surface will also give lower carbon deposition rates, $X_c N_G$, than the stone circulation method, because $N_G = N_{G_{MIN}}$ in these two schemes.

Thus, the gasifier thermal efficiency may vary between 97% and 85% depending on the carbon deposition rate and the choice of gasifier and regenerator temperature control schemes. As the air/fuel ratio approaches the minimum value for gasifier temperature maintenance, all of the proposed gasifier temperature control schemes will tend to approach the behavior of the curve for the minimum air/fuel ratio in Figures 2-10 and 2-11; while as the air/fuel ratio is increased, the proposed temperature control schemes, steam and water injection, will become less attractive.

The projected gasifier product heating value is shown in Figure 2-12, for the regenerative operation, as a function of the carbon deposition rate. Curves are shown for gasifier temperature control by use of heat transfer surface, water injection, steam injection, and stack gas recycle at 20% (solid curves) and 25% of stoichiometric air (broken curves). The product heating value for the case of operation at the minimum air fuel ratio is about 500 Btu/S ft^3. The method of regenerator temperature control is also indicated on the figure, with the use of make-up limestone ($N_G = N_{G_{MIN}}$) giving, in general, slightly higher heating values than the use of stone circulation ($N_{MR} = 0$). The use of regenerator heat transfer surface gives heating values identical to the case of regenerator make-up limestone.

Curve 643984-B

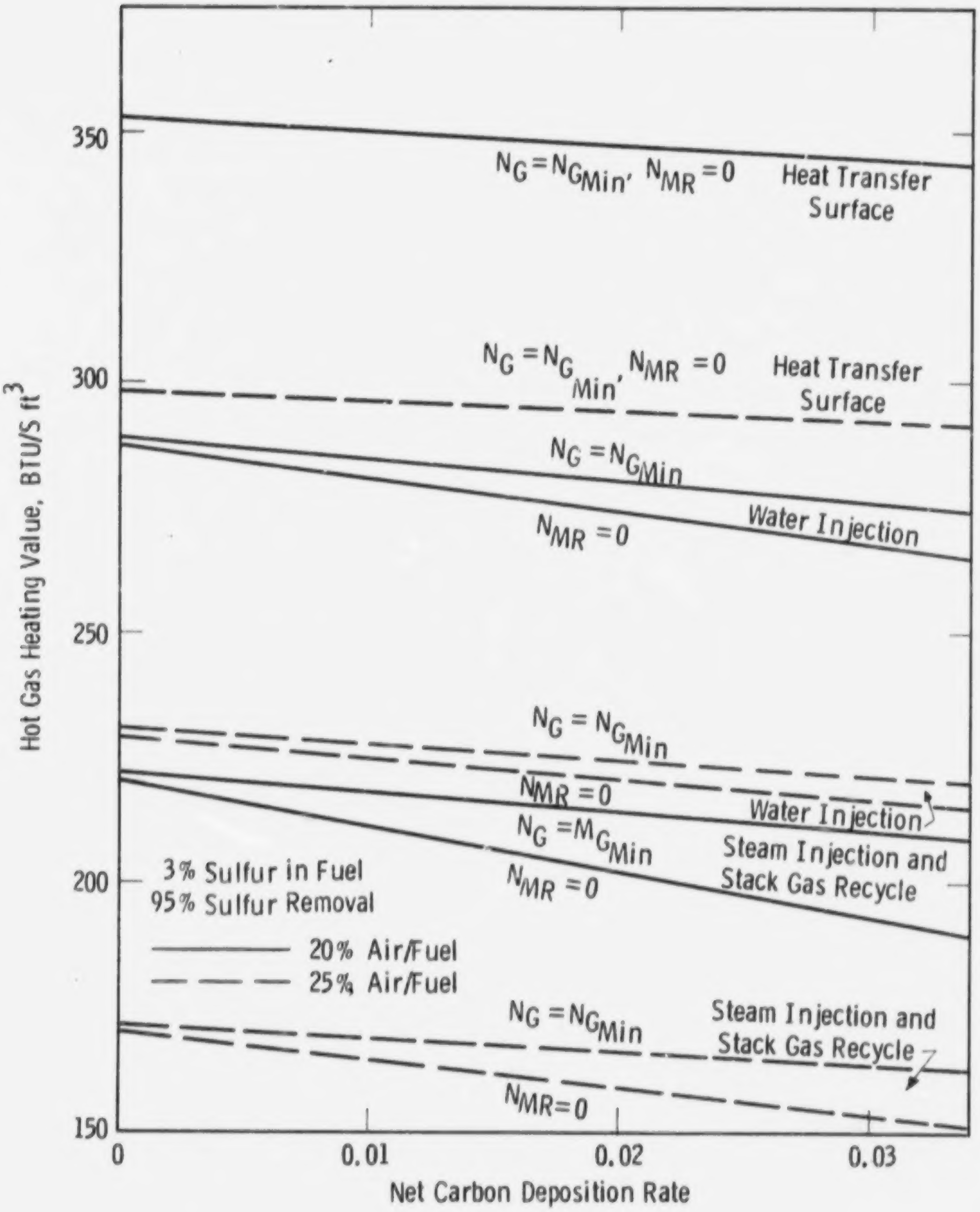

Fig. 2-12—Gasifier Product Heating Value

The overall behavior indicated in Figure 2-12 shows that the gasifier product hot gas heating value is relatively insensitive to the carbon deposition rate and the method of regenerator temperature control, but is largely dependent on the air/fuel ratio and the gasifier temperature control method. The product heating value is an indication of the relative equipment sizes necessary with the various temperature control schemes.

The next matrix summarizes the advantages and disadvantages of the five proposed gasifier temperature control schemes based on the considerations made in this section. The low thermal efficiency of steam injection at air/fuel ratios of 20% to 25% of stoichiometric makes control by steam injection unattractive, while the large equipment sizes required with stack gas recycle at air/fuel ratios of 20% to 25% of stoichiometric make stack gas recycle unattractive. As the air/fuel ratio is reduced, all of the proposed gasifier temperature control schemes become more attractive; with operation of the gasifier at the minimum air/fuel ratio of about 14% of stoichiometric the most attractive condition, if the carbon deposition rate does not increase to unreasonable values (X_c approaches 0.01). Economic analysis of the regenerative operation adds some quantitative measures as a basis to compare the proposed temperature control schemes.

Once-Through Operation

In the once-through gasification/desulfurization operation, the same five possible gasifier temperature control methods are examined. The sulfate generator temperature of 1500°F is maintained by one of two proposed methods: heat transfer surface in the sulfate generator or circulation of excess air through the sulfate generator. The overall behavior of the once-through operation differs from that of the regenerative operation because of the higher rate of limestone addition to the gasifier in the once-through case and the nature of the sulfate generator.

EVALUATION OF GASIFIER TEMPERATURE CONTROL METHODS

CONTROL METHOD	ADVANTAGES	DISADVANTAGES
Heat Transfer Surface	Control is independent of the process operation Limits equipment sizes, HV ≈ 325 Btu/S ft$^{3(a)}$ Maximizes thermal efficiency ≈ 97%	High capital costs Complex design and control Probably necessitates heat transfer surface in the bed, or modular design
Minimum Air/Fuel Ratio	Permits operation without further temperature control provisions in gasifier Minimizes equipment sizes HV ≈ 500 Btu/S ft^3 Maximizes thermal efficiency ≈ 97%	Possible high carbon deposition rate
Stack Gas Recycle	Simple procedure High thermal efficiency ≈ 97%	Necessitates large equipment sizes, HV ≈ 200 Btu/S ft^{2}(a) Highest power requirements, particle clean-up costs
Steam Injection	Simple gasifier design May lower carbon deposition rate	Low thermal efficiency ≈ 93%$^{(a)}$ Necessitates large equipment sizes, HV ≈ 200 Btu/S ft^{2}(a) May be detrimental to sulfur removal
Water Injection	Simple gasifier design May lower carbon deposition rate Fuel distribution may be improved	Fairly low thermal efficiency ≈ 93%$^{(a)}$ Water distribution may be a problem area HV ≈ 250 Btu/S ft$^{3(a)}$ May be detrimental to sulfur removal

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(a) At an air/fuel ratio of 20% to 25% of stoichiometric.

Sulfate Generator

Heat is produced in the sulfate generator by the same reactions operable in the regenerator, equations (2,9), (2-10), and (2-11), at the temperature of 1500°F, reaction (2-10) should dominate while complete combustion of the carbon deposited on the gasifier lime is assumed. The rate of heat generation by chemical reaction, ΔH_S, in the sulfate generator is assumed to be

$$\Delta H_S = 14{,}093\, X_c N_G + 3.96 \times 10^5 \frac{Y\varepsilon_s}{32} \qquad (2\text{-}25)$$

in terms of the carbon deposition rate, $X_c N_G$, and the rate of sulfur removal, $Y\varepsilon_s$. Utilized lime enters the sulfate generator at a rate N_G and a temperature of 1600°F, leaving at a temperature of 1500°F. If E is the fraction of excess air circulated through the sulfate generator then an energy balance around the sulfate generator gives, for temperature control,

$$0 = \Delta H_S + 30\, N_G\, (1 + E) \left[\frac{X_c N_G}{12} + \frac{Y\varepsilon_s}{32}\right] \left[\frac{0.79}{0.21}\, 7.4\, (1423)\right]$$
$$-11.6 \left(\frac{X_c N_G}{12}\right) 1423 - 7.8 \left(\frac{X_c N_G}{12} + 2\, \frac{Y\varepsilon_s}{32}\right) E\, (1423) \quad , \qquad (2\text{-}26)$$

assuming no heat losses. This reduces to

$$E = \frac{X_c N_G + 1.05\, Y\varepsilon_s}{0.448\, X_c N_G + 0.337\, Y\varepsilon_s} \quad . \qquad (2\text{-}27)$$

Note that in the once-through operation N_G is much smaller than in the regenerative case. Where N_G may be as large as 2 to 3 lb/hr/lb fuel/hr in the regenerative operation, N_G will be about 0.2 lb/hr/lb fuel/hr in the once-through case, making $X_c N_G << Y\varepsilon_s$. Thus, carbon deposition may be neglected in equation (2-27) to give

$$E \simeq 3.1 \;,$$

or 310% excess air must be circulated through the sulfate generator to maintain temperature control, independent of the rate of sulfur removal from the residual fuel oil.

It appears that operating with excess air in the sulfate generator should promote a high production of calcium sulfate, so that excess air should be used even in the case of temperature control of the sulfate generator with heat transfer surface. At a superficial velocity of 8 ft/sec, with 95% removal from a 3 wt % sulfur fuel oil, and a 95% thermal efficiency of the once-through operation, the cross-sectional area of the sulfate generator would be 0.285 ft^2/MW with temperature control by heat transfer surface (based on 50% excess air). With temperature control by excess air circulation the cross-sectional area would be about 1.0 ft^2/MW. The cross-section of the sulfate generator with control by heat transfer surface is nearly the same as that of the regenerator in the regenerative operation (Figure 2-8). However, with the rate of heat generation about three times as great in the sulfate generator as in the regenerator, much more heat transfer surface will be required for temperature control. Design will probably be limited to placing heat transfer surface internal to the sulfate generator if this method of temperature control is to be used.

The temperature control scheme using excess air circulation through the sulfate generator is clearly superior to heat transfer surface: capital costs should be lower, and the operation and control simpler.

Gasifier - Once-Through Operation

Stack gas recycle, steam injection, water injection, heat transfer surface, and operation at the minimum air/fuel ratio are proposed for temperature control of the once-through gasifier. An energy balance around the gasifier is made up of the following factors: with a limestone make-up rate to the gasifier of 3 moles CaO/mole sulfur fed, the heat taken up in the calcination of the make-up limestone is three times that assumed in the regenerative case, or

Energy to calcination = 1.137×10^4 Y Btu/lb fuel
Energy lost through boundaries = 177 Btu/lb fuel
Energy returned from sulfate generator = ΔH_S.

An energy balance gives

$$F_S \Delta H_S = \Delta H_G + \Delta H_R - 1.137 \times 10^4 Y - 177 + F_P \Delta H_P \qquad (2\text{-}28)$$

in terms of the nomenclature introduced earlier. Equation (2-28) indicates the requirements for temperature control with stack gas recycle, steam injection, and heat transfer surface. Comparing equation (2-28) with the results for the regenerative operation, equation (2-23), F_S is nearly identical in the two operations for equal air/fuel ratio and carbon deposition rate.

Since the heat generated in the sulfate generator, ΔH_S, is recycled to the gasifier, higher thermal efficiencies are expected in the once-through operation than were found for the regenerative operation. In terms of the energy taken from the gasifier product, the thermal efficiency becomes

$$1 - \varepsilon_T = \frac{10{,}651\ Y + 177 - 1750\ \varepsilon_s Y + F_S \lambda}{17{,}700} . \qquad (2\text{-}29)$$

For the gasifier temperature control methods of stack gas recycle, heat transfer surface, and minimum air/fuel ratio, the thermal efficiency is independent of the carbon deposition rate in contrast to the regenerative case. Nor does the air/fuel ratio affect the thermal efficiency with stack gas recycle and heat transfer surface. For 95% removal of sulfur in a 3 wt % sulfur fuel, the thermal efficiency is 97.5% with stack gas recycle, heat transfer surface, and minimum air/fuel ratio. For steam and water injection the thermal efficiency is reduced by the factor $F_S \lambda$ much as in the regenerative case.

The heating value of the gasifier product from the once-through operation is nearly the same as the heating value obtained for the regenerative operation. As in the regenerative operation, the heating value of the once-through gasifier product is sensitive to the air/fuel ratio and the gasifier temperature control method, but not to the rate of carbon deposition or the method of sulfate generator temperature control.

In the once-through case, operation of the gasifier at the minimum air/fuel ratio appears superior to the gasifier temperature control schemes proposed for use at higher air/fuel ratios. Large fractions of deposited carbon on the gasifier lime, X_c, will be more tolerable in the once-through operation than in the regenerative operation unless the rate of sulfur removal is reduced by the existence of carbon in large amounts. This behavior may make operation at the minimum air/fuel ratio more easily realized in the once-through operation than in the regenerative operation.

Overall Material and Energy Balances

Based on the specifications presented on page 461, material and energy balances have been carried out for regenerative and once-through operations. Results are listed in Tables 2-5 and 2-6 for the two schemes whose flow diagrams are shown in Figures 2-13 and 2-14.

For the regenerative operation, Table 2-5 indicates the stream characteristics for the five proposed gasifier temperature control schemes and the three regenerator temperature control schemes. In addition to mass flow rates and temperatures of the streams shown in Figure 2-13, Table 2-5 presents other pertinent characteristics of the gasifier product, the cross-sectional area of the gasifier and regenerator per lb of fuel/hr, and the estimated bed depths of the gasifier and regenerator.

For the once-through operation pictured in Figure 2-14, Table 2-6 indicates the stream characteristics and vessel dimensions for the gasifier control schemes of stack gas recycle, minimum air/fuel ratio, and heat transfer surface, and for sulfate generator temperature control by excess air circulation. There is general similarity between the regenerative and once-through operations, with the greatest difference in the two operations in the stone flow rates.

The importance of the gasifier temperature control scheme is evident from Tables 2-5 and 2-6. The equipment sizes, thermal efficiency, and stone circulation rates vary greatly depending on the choice of temperature control.

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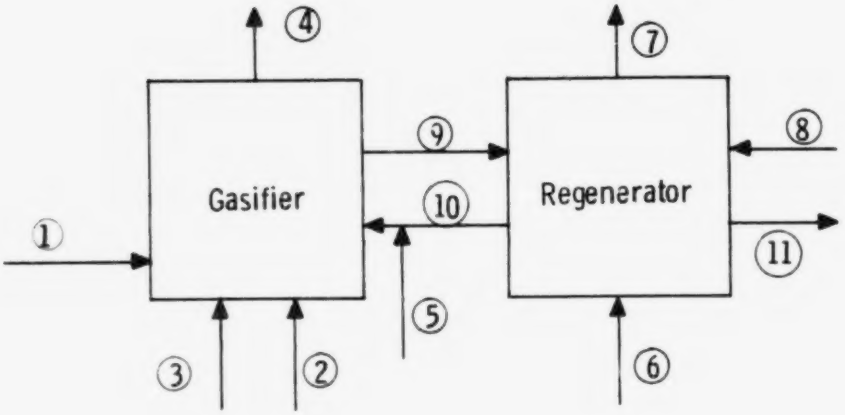


Fig. 2-13—Regenerative Operation Flow Diagram

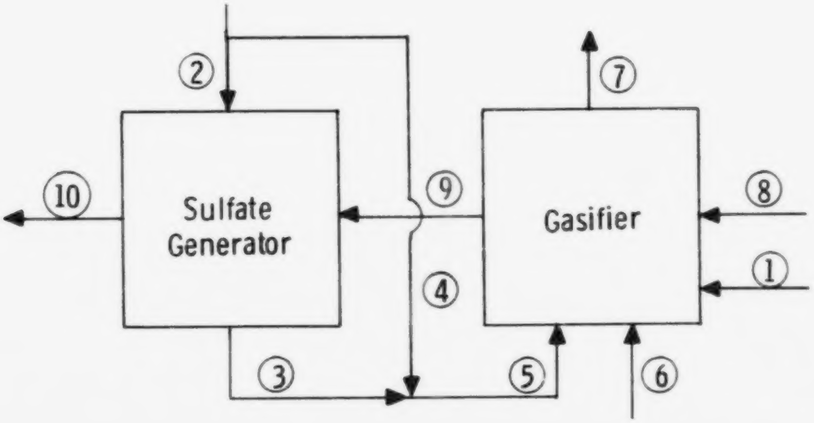


Fig. 2-14—Once-Through Operation Flow Diagram

TABLE 2-5

REGENERATIVE OPERATION MATERIAL AND ENERGY BALANCES

STEAM	TEMP.	GASIFIER TEMPERATURE CONTROL METHOD				
		Stack Gas Recycle	Minimum Air/Fuel	Heat Transfer Surface	Steam Injection	Water Injection
Fuel oil, lb/hr	200°F	1	1	1	1	1
Air, lb/hr	77°F	2.76(a)	1.94(b)	2.76	2.76	2.76
Temperature control stream, lb/hr(c)		1.70 [1.50](d)	0	~ 850 Btu/hr Heat transfer rate	0.93 [0.78]	0.37 [0.30]
Gasifier product, lb/hr	1600°F	5.46	2.94	3.76	4.69	4.13
ft^3/hr		320	138	195	299	236
Btu/S ft^3		215	498	352	212	283
ϵ_T(e)		97.1	97.1	97.1	92.2	95.3
Gasifier make-up, lb/hr	77°F	0.095	0.095	0.095	0.095	0.095
Gasifier area, ft^2/lb fuel/hr(f)	1600°F	0.0093	0.003	0.0053	0.009	0.0064
Static depth, ft		2.5-3.5				

STEAM	TEMP.	REGENERATOR TEMPERATURE CONTROL METHOD		
		Stone Circulation	Make-Up Stone	Heat Transfer Surface
Regenerator air, lb/hr	77°F	0.302	0.228	0.228
Regenerator product, lb/hr	1900°F	0.34	0.268	0.268
X_{SO_2}		0.09	0.12	0.12
Regenerator make-up, lb/hr	77°F	0	0.049	0
Stone circulation, lb/hr	1600°F	1.80	0.713	0.713
$X_C N_G$		0.01	0.003	0.003
Stone circulation, lb/hr	1900°F	1.75	0.68	0.66
Sulfated stone disposal, lb/hr	1900°F	0.051	0.077	0.051
Regenerator area, ft^2/lb fuel/hr(f)	1900°F	6.23×10^{-4}	4.7×10^{-4}	4.7×10^{-4}
Static depth, ft		2.0		

(a) 20% of stoichiometric
(b) 14% of stoichiometric
(c) Based on 300°F stack gas, 212°F steam
(d) Figures in brackets refer to regenerator temp. control by make-up stone or H.T.S.
(e) Regenerator temperature control by stone circulation
(f) Based on superficial velocity of 8 ft/sec

TABLE 2-6

ONCE-THROUGH OPERATION MATERIAL AND ENERGY BALANCES

STREAM	Temp.	GASIFIER TEMPERATURE CONTROL METHOD		
		Stack Gas Recycle	Minimum Air/Fuel	Heat Transfer Surface
Fuel oil, lb/hr	200°F	1	1	1
Sulfate generator air, lb/hr	77°F	0.983[a]	0.983	0.983
Sulfate generator product, lb/hr	1500°F	0.924	0.924	0.924
lb O_2/hr		0.171[b]	0.171	0.171
Air, lb/hr	77°F	2.02	1.15	2.02
Gasifier air stream, lb/hr		2.94 [490°F]	2.07 [790°F]	2.94 [490°F]
Temperature control stream, lb/hr		1.60 [300°F]	0	~ 650 Btu/lb fuel Heat transfer rate
Gasifier product, lb/hr	1600°F	5.54	2.85	3.94
ft^3/hr		296	148	204
Btu/S ft^3		231	465	336
ε_T		0.975	0.975	0.975
Limestone feed, lb/hr	77°F	0.265	0.265	0.265
Sulfided lime, lb/hr	1600°F	0.180	0.180	0.180
Sulfated lime, lb/hr	1500°F	0.186	0.186	0.186
Gasifier area, ft^2/lb fuel/hr	1600°F	0.0085	0.0033	0.0053
Static depth, ft		3.5-4.0	3.5-4.0	3.5-4.0
Sulfate generator area, ft^2/lb fuel/hr		0.0017	0.0017	0.0017
Static depth, ft		3-4	3-4	3-4

[a] Based on 310% excess air

[b] < 0.1% SO_2 in product

Power Requirements

Pressure drops have been estimated in Table 2-7 for regenerative and once-through operations in which low-pressure drop burners requiring no particulate clean-up prior to combustion have been used. Bed pressure drops are estimated to be about one inch of H_2O per inch of bed height; distributor pressure drops are assumed to be 50% of the bed ΔP; and burners are assumed to contribute two to four inches of H_2O pressure drop. In the case of the sulfate generator, particulate removal has been assumed necessary in the recycle stream going from the sulfate generator to the gasifier to protect the air distributor. In addition, the air-side pressure drop in the burners is assumed to be 7 to 20 inches of H_2O.

Power requirements, exclusive of power for solids handling and circulation, are listed in Table 2-8 and are based on the maximum pressure drops shown in Table 2-7. The table indicates the power requirements per pound of fuel over and above that already required in the power generating system. This power requirement is combined in Table 2-8 with the thermal efficiency of the gasifier to give the overall efficiency of the gasification/desulfurization system. Though the power requirements of the once-through operation are somewhat larger than the power requirements of the regenerative system, the once-through operation has a higher overall efficiency. Auxiliary power costs do not appear to be an extremely important operating cost factor with gasification systems, totalling at most a loss of about 1% in overall efficiency.

TABLE 2-7

PRESSURE DROPS

	REGENERATIVE OPERATION 8 ft/sec		ONCE-THROUGH OPERATION 8 ft/sec	
	Gasifier (in. H_2O)	Regenerator	Gasifier	Sulfate Generator
Bed	30-42	24	42-48	36-48
Distributor	15-21	12	21-24	18-24
Piping	5-7	4	7-8	7-10
Burners	2-4	---	2-4	---
Cyclone	---	---	---	20-30
TOTAL	52-74	40	72-84	81-112

TABLE 2-8

POWER REQUIREMENTS

	REGENERATIVE OPERATION					ONCE-THROUGH OPERATION		
	Stack Gas Recycle	Minimum Air/Fuel	Heat Transfer Surface	Steam Injection	Water Injection	Stack Gas Recycle	Minimum Air/Fuel	Heat Transfer Surface
Kwh/lb fuel(a)	0.021	0.008	0.013	0.021	0.015	0.022	0.012	0.016
Thermal efficiency	0.968	0.968	0.968	0.902	0.943	0.975	0.975	0.975
Overall efficiency	0.957	0.963	0.960	0.889	0.935	0.960	0.967	0.963

(a) Based on a plant heat rate of 10^4 Btu/kwh, a fuel oil heating value of 17,700 Btu/lb, and a compressor efficiency of 70%.

DESIGN CONCEPTS

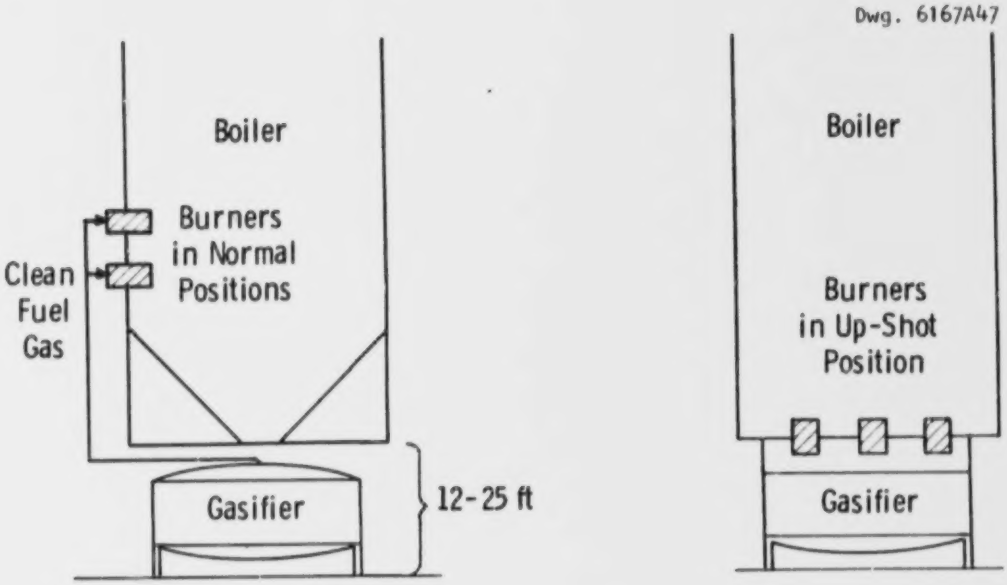
Energy and material balances provide basic information with which to examine the feasibility of applying the gasification/desulfurization concept as an add-on to an existing boiler, or as a new plant design feature. The feasibility of the retrofit concept is investigated in terms of the availability of space in an existing power plant, the modifications necessary to retrofit an existing boiler, and the performance of a modified boiler. The performance and design concepts projected for new boiler installations are examined.

Retrofit Concepts

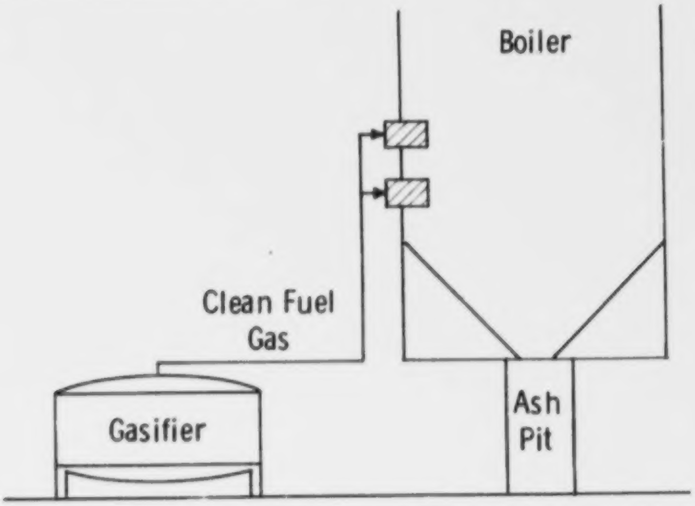
The feasibility of converting an existing boiler to one which utilizes the fuel from a gasification/desulfurization system depends on a number of factors, many of which will differ from one boiler to the next. The space available for a gasification/desulfurization system in an existing power plant, the modifications necessary to retrofit an existing boiler, and the performance of a retrofit boiler will depend on the specific gasification/desulfurization system design and choice of operating conditions, the location of the gasification/desulfurization system in the plant, the type (coal- or oil-fired) of boiler, size of boiler, the turn-down needed, the boiler load factor, and the specific design features of the boiler. These factors are discussed with respect to the location of the gasification/desulfurization system relative to the boiler, and the performance of the retrofit boiler.

Retrofit Location

The gasification system may be placed internal to the boiler (directly beneath) or external to the boiler (as close as possible without carrying out major boiler modifications in order to locate the system). These concepts are illustrated in Figure 2-15. The feasibility of the internal location is examined in Figures 2-16 and 2-17 with respect to the amount of space available beneath the boiler. The two figures compare the cross-sectional area of coal- and oil-fired boilers with the MW rating of the boiler. Superimposed on the figure are the gasifier cross-sectional areas for the various temperature control schemes plotted against the gasifier MW rating. Figure 2-16 is for an air/fuel ratio of 20% of stoichiometric, while Figure 2-17 is at 25% of stoichiometric, both figures are at a superficial velocity of 8 ft/sec. In interpreting the figures, it should be noted that the regenerator, or sulfate generator cross-section is much smaller than



Internal Gasifier Location



External Gasifier Location

Fig. 2-15—Boiler retrofit concepts

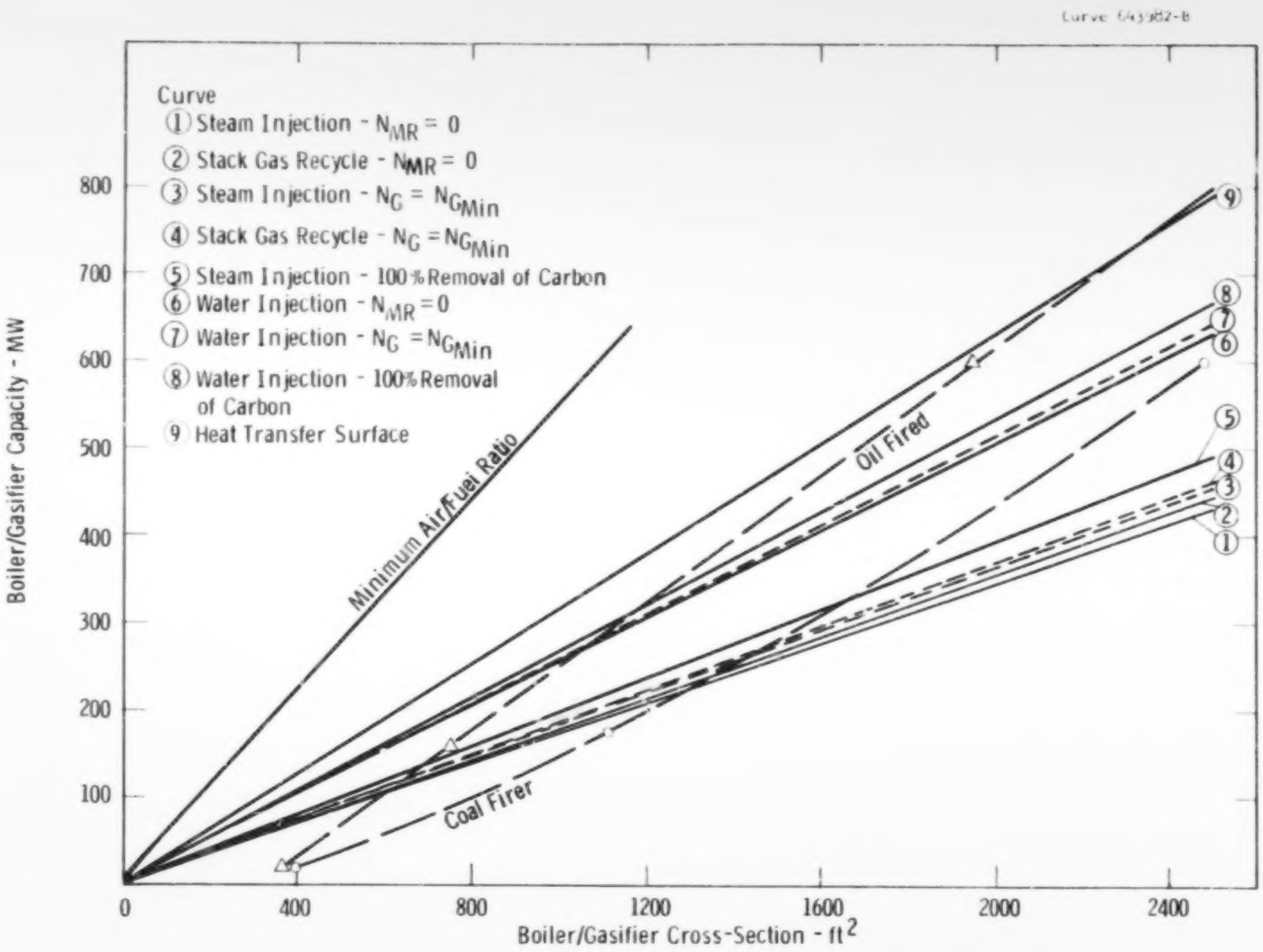


Fig. 2-16—Space Requirements for Internal Design - 20% Stoichiometric Air: 8ft/sec - Carbon Deposition Rate = 0.015 lb Carbon/lb Fuel

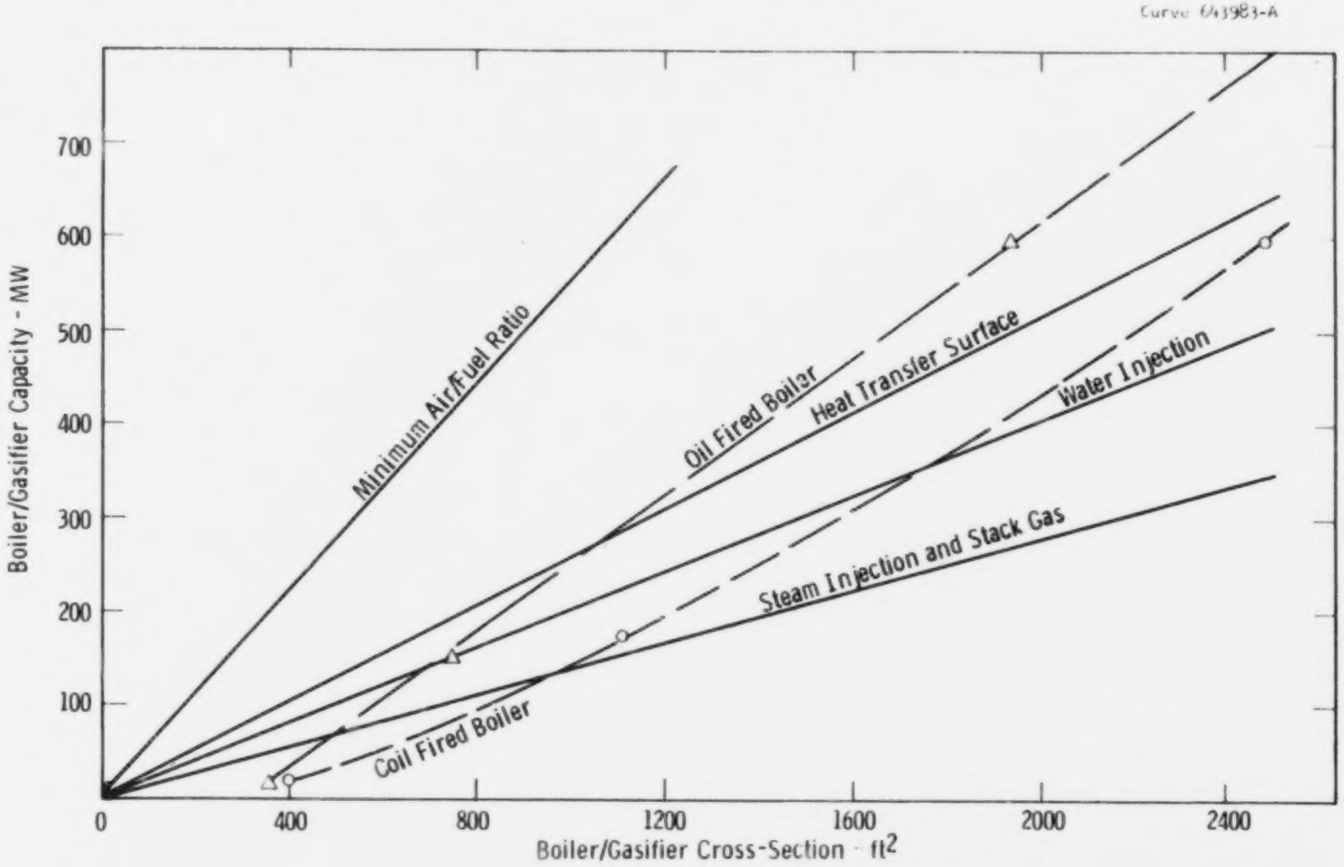


Fig. 2-17– Space Requirements for Internal Design, 25% Stoichiometric Air: 8ft/sec - Carbon Deposition Rate = 0.004 lb Carbon/lb Fuel

the gasifier cross-section and was not included in the figures. With respect to vertical space, 12 to 25 ft of headroom is available under boilers, which is sufficient for a gasifier 10 ft in total height.

The criterion that the internal design is feasible if the boiler cross-section is larger than the gasifier cross-section suggests the following retrofit design factors:

Critical Factors in Determining Feasibility of Internal Retrofit Design--Regenerative or Once-Through:

- Superficial velocity
- Air/fuel ratio
- Temperature control method
- Boiler type (coal or oil) and size

Feasibility Limits on the Internal Retrofit (Based on Boiler Capacity > 100 MW):

- Superficial velocity $\geq$ 8 ft/sec
- Air/fuel ratio $\leq$ 20% of stoichiometric
- Temperature control by minimum air/fuel ratio and heat transfer surface feasible for boilers > 100 MW
- Temperature control by stack gas recycle, steam and water injection feasible for boilers < 200 MW
- Internal design more probable with coal- than with oil-fired boilers.

The curve showing the cross-sectional area of the gasifier with temperature control by operation at the minimum air/fuel ratio represents the minimum cross-sectional area possible at a given superficial velocity and is the most promising operating condition from the standpoint of the internal design concept. The gasifier cross-section curves in Figures 2-16 and 2-17 refer specifically to the regenerative operation, but they are also representative of the once-through operation gasifier cross-sections.

Another factor which may limit the feasibility of the internal design is the need for particulate removal before the gasifier product reaches the burners. Though preliminary investigation indicates that burners may be designed which can be operated at high dust loading levels, the situation where high-efficiency cyclones are called for is illustrated in Figures 2-18 and 2-19. Figure 2-18 shows a 20 MW coal- or oil-fired boiler with a regenerative oil gasifier add-on operated at 6 ft/sec, 20% of the stoichiometric air/fuel ratio, with temperature control by water injection, and burners modified to be in the upshot position. Figure 2-19 shows a 150 MW coal-fired boiler retrofit with a regenerative gasification system with the same operating conditions as in Figure 2-18. In this case the burners have been left in the normal boiler positions, and high-efficiency cyclones have been placed next to the boiler due to their excessive height.

External location of the gasification/desulfurization system would probably minimize boiler modifications, though it might necessitate the removal of auxiliary equipment in the proximity of the boiler. The greatest concern with the external design is the necessity of transporting the 1600°F reducing gas from the gasifier to the boiler. Great lengths of piping could be expensive, and heat losses and condensations of tars in the piping must be minimized. The internal design concept may exhibit the same problem. Figure 2-20 illustrates a modular external design for a 600 MW coal-fired boiler. High-efficiency cyclones are again included.

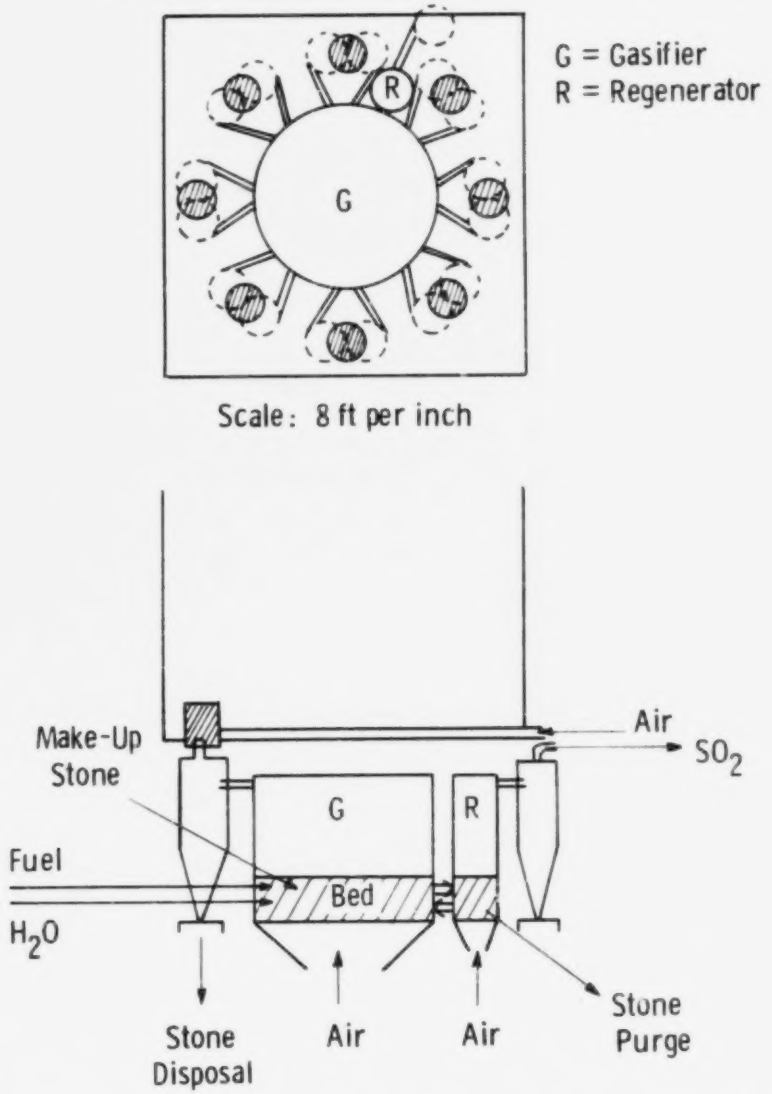
A more detailed look at specific boiler designs, and the modifications needed for internal or external retrofit designs, is needed in order to evaluate these concepts. The cost involved in boiler modification and the down-time of the boiler during modification may be the limiting factor in retrofit feasibility.

Boiler Performance and Boiler Modifications

Degraded performance of a retrofit boiler due to a decrease in the temperature and luminosity of the flame produced at the burner may

Dwg. 2950A72

8 - 2.5 × 10 Btu/hr Burners
2 High Efficiency Cyclones per Burner - 3×10^3 ft^3/min per Cyclone



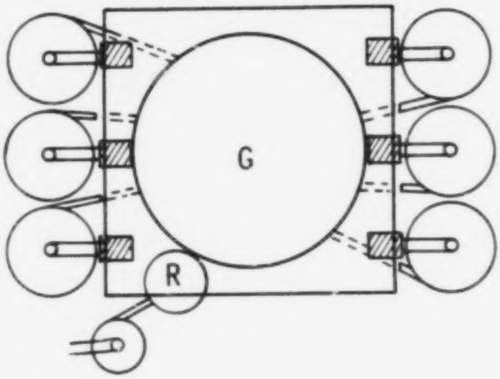
G = Gasifier
R = Regenerator

Scale: 8 ft per inch

Fig. 2-18—20 MW gasifier add-on

Dwg. 2950A71

$6 - 2.5 \times 10^8$ Btu/hr Burners
1 High Efficiency Cyclone per Burner



Scale: 16 ft per inch

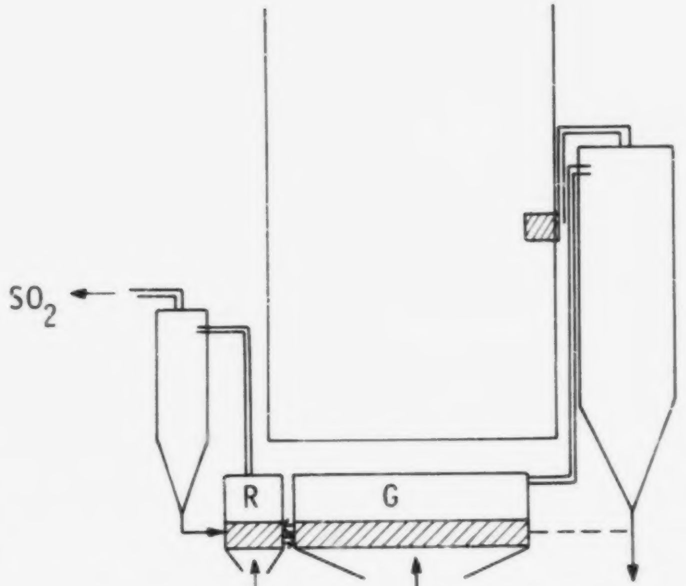


Fig. 2-19—150 MW gasifier add-on to a coal fired boiler

Dwg. 2950A73

4 - 150 MW Modules; 6 high efficiency cyclones per Module

24 - 2.5 × 10^8 Btu/hr Burners

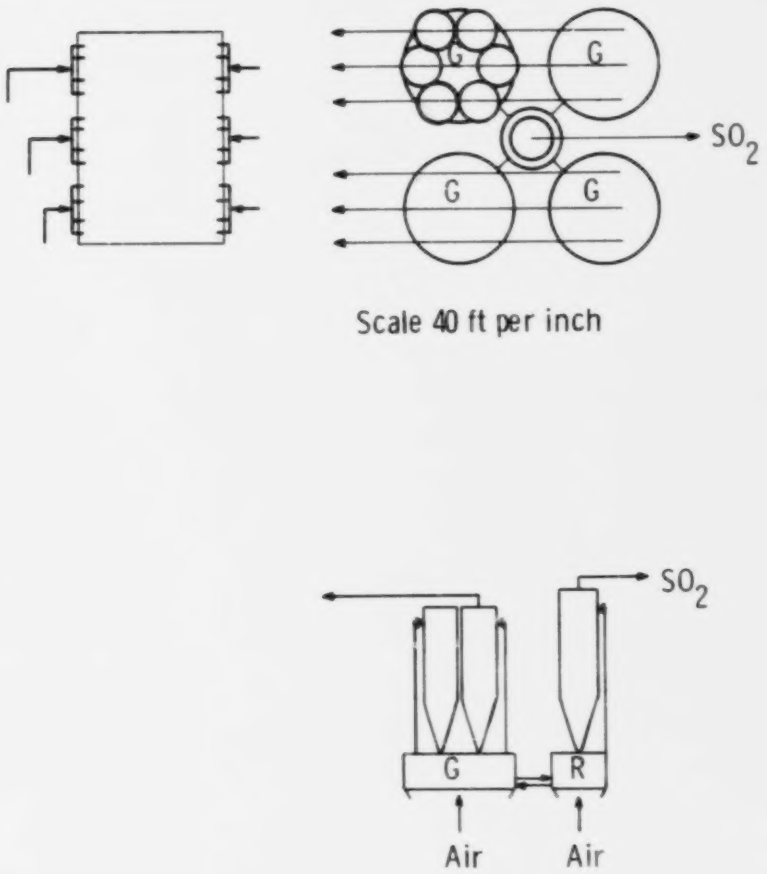


Fig. 2-20—600 MW gasifier external to a coal fired boiler

be a problem to be considered with all retrofit boilers. The main factors controlling the degradation of boiler performance are the gasifier temperature control method and the boiler type. If no inert material is added to the gasification product to control temperature, and if the thermal efficiency of the gasifier is high, then the boiler degradation should be mainly a function of the flame luminosity. Esso (England) has obtained a luminous flame upon combustion of the gasification product, indicating that the temperature control scheme may be the most important single factor. Note that a decrease in the flame temperature due to the addition of a noncombustible to the fuel will be accompanied by an increase in gas velocity through the convective section of the boiler, yielding higher heat transfer in that section.

The time and capital involved in the modification of a utility boiler are important factors affecting the feasibility of boiler retrofits. Any retrofit, of external or internal location, calls for the replacement of the boiler burners. Burner modifications will include modifications to the burner manifolding, alterations to the boiler water walls, and possibly changes in the structural foundation of the boiler walls. Minimizing the new burner size will be advantageous from this standpoint, since the size is mainly a function of the gasifier temperature control scheme and the air/fuel ratio.

The internal add-on concept calls for the removal of the boiler ash pit and other auxiliary equipment found beneath the boiler. Alterations to the lower sloping water walls of the boiler will be necessary if the burners are placed in the upshot position at the base of the boiler. Further modifications result from the need for additional fans and compressors, and particulate clean-up equipment.

The following two matrices summarize the points affecting the feasibility of boiler retrofits. The first compares the internal and external location of the gasification/desulfurization system, while the second compares the modifications and boiler performance probable with coal- and oil-fired retrofits.

The factors presented in these matrices favor the external retrofit design over the internal retrofit design, and the feasibility of retrofitting a coal-fired boiler over the feasibility of retrofitting an oil-fired boiler. However, these preliminary investigations do not eliminate the possibility of the internal retrofit design or the retrofit of oil-fired boilers.

GASIFIER RETROFIT LOCATION FACTORS

Internal Location	External Location
Space Requirements	
Under suitable operating conditions space available beneath the boiler may be utilized (page 500)	Space must be found or provided as near to boiler as possible
Cross-sectional area needed to place burners in the upshot position at base of boiler may limit concept (new burners may be 1.5-3 times the size of normal oil or coal burners)	More flexibility in the external design may allow operating with maximum fuel heating value to minimize burner sizes
Design limited by available overhead space if particulate control is needed before burners	Feasibility not as limited by space required for modular design, overhead space, or other spatial considerations
Probably not feasible with modular design in large boilers (> 500 MW) except at velocities >> 8 ft/sec	
Modification Factor	
More extensive modifications required, more down time involved in removal of ash pit, auxiliary equipment, alterations to water walls and foundations	Modifications consist of removal of auxiliary equipment, such as pulverizers, and addition of new burners
Construction more expensive in confined quarters beneath boiler	Prefabrication of vessels possible
Piping and manifolding may be minimized with burners placed in base of boiler	Relatively great amount of high-temperature, noncorrosive piping and manifolding necessary, though the same may be true with internal design
Each boiler retrofit design may need to be treated as a unique case	More uniformity in design from one retrofit to another
Performance Factor	
Need for operating conditions giving maximum gasifier compactness may be detrimental to system performance	Greater flexibility in operating conditions with external design may allow operation at optimum conditions

RETROFIT OF COAL- AND OIL-FIRED BOILERS

Space Requirements

Coal-fired boilers have larger cross-sections than oil-fired boilers at large capacity (> 100 MW)

Cross-sections are comparable for smaller capacity boilers (< 100 MW)

Headroom is nearly the same in coal- and oil-fired boilers at 12 to 25 feet

More potential space exists near a coal-fired boiler than near an oil-fired boiler - pulverizers and other auxiliary equipment may be removed - especially advantageous if the coal boiler has been previously converted to operate with low-sulfur oil.

Modification Factors

Normal coal and oil burners are nearly the same size (200 MW Btu/hr burner is 10 to 11 ft^2 in port cross-section)

Much solid handling equipment and particulate clean-up equipment may already exist with a coal-fired boiler. Increased particulate control would probably be needed with an oil-fired boiler.

Performance Factors

Coal-fired boilers are designed to operate with slag on the boiler heat transfer surface, which may make degradation of boiler performance negligible with coal-fired boilers. Boiler performance degradation may be important with oil-fired boilers depending on the gasifier temperature control method.

New Boilers

In contrast to boiler retrofit considerations, the feasibility of incorporating a gasification/desulfurization system into a new boiler design will be limited only by the overall economics of the system and the market potential for new oil-fired boilers. The total space occupied by the gasification/desulfurization system will be a small percentage of the total plant volume. Also, because of the flexibility in boiler design, the boiler performance will not be affected by the presence of the gasification/desulfurization system.

Market studies indicate a decrease in the number of new oil-fired boilers which will be constructed, with practically no new oil-fired boilers being constructed after 1980.(1) Based on this projection, the gasification/desulfurization of residual oil will be limited to the retrofit of existing boilers, regardless of the system economics. Further investigation of the market potential of the new boiler system is needed.

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(1) L. G. Hauser, private communication.

Burners

Discussions with the Bloom Engineering Company and Process Combustion Corporation suggest that with little or no development work, burners can be constructed which can handle the hot gasifier product, combust with low excess air to give a stable flame, and simultaneously operate with high dust loading to eliminate the need for cyclones prior to combustion. Bloom Engineering has had considerable experience with boiler burners handling hot, dirty, low-Btu fuels such as lean coke oven gas and blast furnace gas, and they have made the following points:

H_2S Corrosion. H_2S corrosion can be prevented by proper choice of materials - stainless steel or possibly refractory.

Dust Loading. Though Bloom has had no experience with dust loading as high as might be expected with a gasification-desulfurization system, and no direct experience with limestone, they felt erosion would not be a problem if the gas fuel were carried by the large central burner port. The burners could be fired in the upshot position with few problems due to limestone deposition and minimum maintenance. The effect of the limestone on the flame temperature, even at 5 grains/S ft^3, will be negligible.

Pressure Drop. About 2 inches of H_2O pressure drop is expected across the central gas inlet, and 7 to 20 inches H_2O across the air inlet, though it may be desirable to increase the pressure drop across the gas inlet for control purposes.

Control-Turn-down. Turn-down with flame stability could be maintained at greater than a 5:1 turn-down ratio. Control of burner intakes could be handled with a distribution orifice at each gas inlet to eliminate manifolding in the case of the internal gasifier location.

Cost. An estimated cost of ∼ \$12,000 was placed on a 100×10^6 Btu/hr burner and port block with stainless steel construction once a commercial design has been established.

Burner Size and Design. Bloom has constructed burners of capacity up to 100×10^6 Btu/hr, and they indicated that larger capacity burners could easily be designed and constructed. Since burner cross-sectional area is approximately inversely proportional to the heating value of the fuel, the heating value should be maximized in order to minimize the boiler modifications required. A preliminary design of a burner is shown in Figure 2-21 to indicate the burner layout and approximate dimensions.

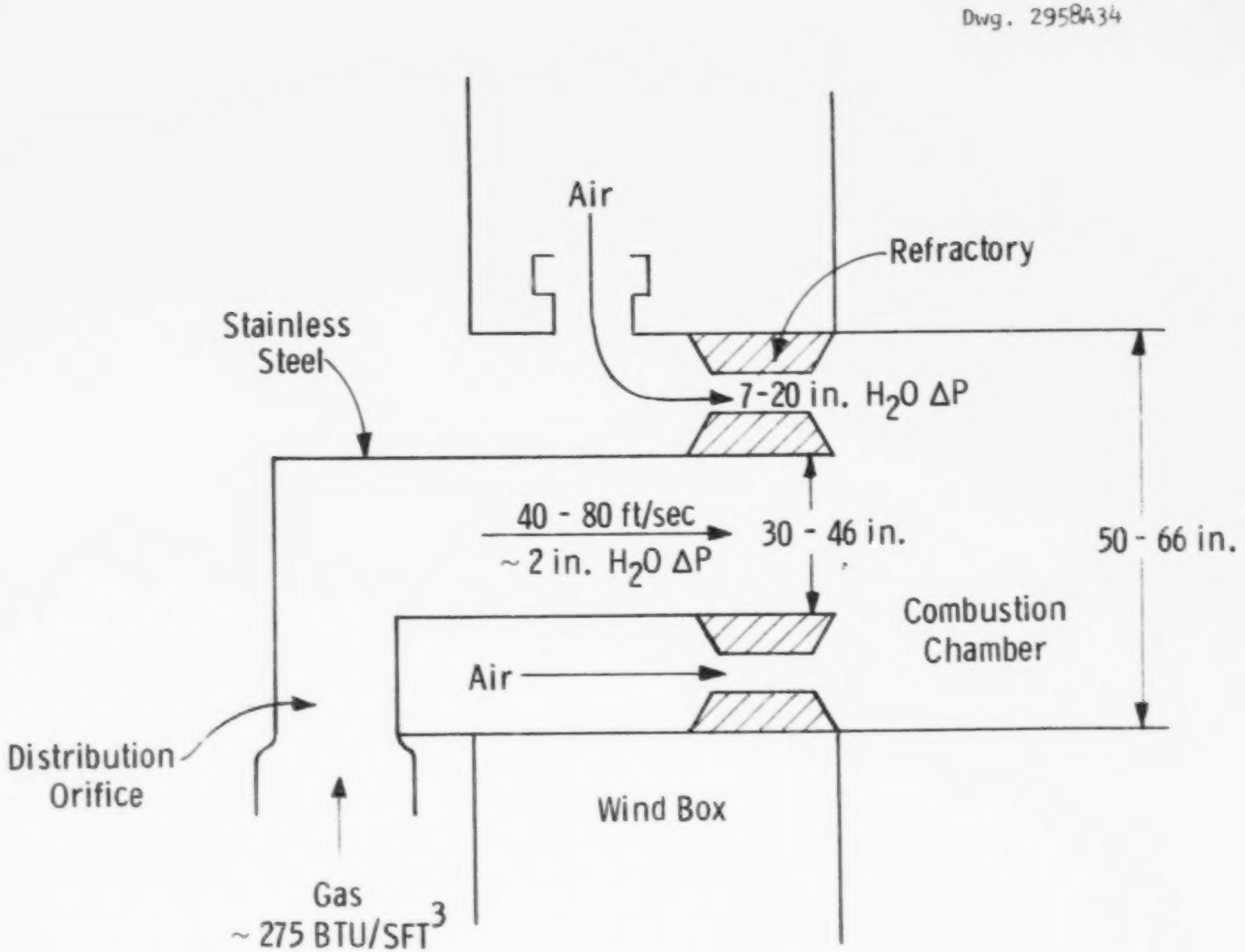

Fig. 2-21—Design of ~2.5 × 10^8 BTU/Hr Burner

Turn-Down

It is expected that gasification/desulfurization systems will be operated at turn-down ratios of up to 4:1. The performance of the gasification/desulfurization system, in terms of fuel properties, sulfur removal, and temperature control, must be considered over a large range of fuel feed rates.

With the system design based on the total boiler capacity, the gasifier performance must continue at a superficial velocity which is 1/4 of the design velocity when operating at maximum turn-down. The critical factor seems to be that the quality of fluidization remain high over this wide range of superficial gasifier velocities. Thus, the maximum operating velocity must be about eight times the minimum fluidization velocity of the lime particle size distribution in the gasifier. The same restrictions hold for the velocities used in the regenerator and sulfate generator. Then, if fluidization with good mixing is maintained at maximum turn-down, the sulfur removal efficiency should increase as the system is turned down due to increased gas residence times in the gasifier; the fuel properties and the temperature control of the system should be unaffected by the turn-down; and the overall operation, including particle carry-over and regeneration or sulfate generation, should improve or remain unaffected as turn-down is achieved.

The possibility of using modular gasifier designs may also aid in the maintenance of the system performance during turn-down. In the case of modular design, individual gasification/desulfurization units may be shut down completely to achieve the required turn-down while the remaining units would continue to operate at or near full capacity.

The details of turn-down, start-up, and other system operation procedures will be considered in the detailed design phase.

ECONOMICS

Capital costs and operating costs have been estimated for both regenerative and once-through operations. Also, both new and retrofit designs of a 600 MW boiler are considered in the economic study for the operating conditions and performance factors listed on page 461.

Cost Factors

The following matrix summarizes the capital and operating cost factors considered:

ECONOMIC STUDY BASIS

CAPITAL COSTS

Equipment includes - retrofit case

Regenerative	Once-Through
Gasifier vessel	Gasifier vessel
Regenerator vessel	Sulfate generator vessel
Limestone handling equip.	Limestone handling equip.
Sulfur recovery plant	Fans, compressors
Fans, compressors	Storage
Storage	Particulate clean-up equip.
Particulate clean-up equip.	Burners
Burners	Boiler modification costs
Boiler modification costs	

Installation charged at 50% of equipment costs - includes materials and labor for concrete structures, piping and ductwork, electrical installation, instrumentation, etc.

Engineering, contingency, and construction charged at 30% of the direct cost.

OPERATING COSTS

Capital Charges - 14% of investment

Maintenance - 5% of investment

Overhead - 4% of investment

Materials -

Limestone - \$2.50/ton
Fuel oil - \$3/bbl
Utilities - Charged as fuel adder in overall system efficiency

Disposal - \$1.30/ton of lime

Labor - neglected

Credits - \$15/ton of sulfur when considered

Load factor - 40% and 80% examined

Equipment costs were generated from the following estimates:

Vessel Costs [30] (installed materials costs)

Stainless steel shell - \$1.80/lb
Insulation - \$20/ft^2
25¢/gas inlet button - 750 buttons/ft^2 of grid
\$27/ft^2 grid surface - included bubble break-up by means of vertical rods above the bed

Assuming the units are 15 ft tall, the vessel cost is given by

$$2220\ D + 167\ D^2,$$

where D is the vessel diameter.

Limestone handling costs consist of two contributions, receiving and disposal equipment costs and recirculation costs. These factors are estimated from the report by Esso R&D [30].

Sulfur Recovery Plant. Direct capital cost and operating cost for the recovery of elemental sulfur were taken from a report by Allied Chemical Corporation [2].

Fans. Costs were estimated from information provided by the Sturtevant Division of Westinghouse.

Particulate Clean-Up Equipment. High-temperature cyclone costs, with stainless steel construction, are taken as twice the cost of low-temperature cyclones. The costs of cyclone clean-up prior to the boiler are included in the analysis. Electrostatic precipitator costs are estimated as \$1.5/kw, not including foundations, to meet present particulate standards.

Burners. A 100 MM Btu/hr burner designed to handle the hot reducing gas is assumed to cost \$12,000. Burner costs are scaled according to their cfm capacity, with \$12,000 taken as a base for a fuel with a heating value of $\sim$ 300 Btu/S ft^3. Burner costs are based on

100 MM Btu/hr burners, although larger capacity burners would probably be used - a 250 MM Btu/hr burner is estimated to cost $20,000-$25,000.

Boiler Modifications. The cost involved in boiler modifications is assumed to result mainly from the power production lost during down time. This is taken at a rate of $5000/day per 100 MW, and a period of 30 days down time is assumed reasonable, though this will vary from boiler to boiler and from one gasifier design to another.

Regenerative Operation Economics - New and Retrofit

Table 2-9 lists the preliminary cost breakdown for the major equipment included in a 600 MW regenerative retrofit and a new system. Total investment is indicated for each of the five proposed temperature control methods in $/kw. The cases of boiler retrofit with and without the cost of particulate control equipment (cyclones prior to combustion and electrostatic precipitator final clean-up) are included, along with the estimated cost for the regenerative gasifier/desulfurization system when combined with a new boiler design.

The table indicates that operation with the minimum air/fuel ratio necessary to achieve temperature control is the most promising control method, if such a low air/fuel ratio can be realized in practice. The other temperature control schemes are of comparable capital cost, with steam injection the most expensive.

Table 2-10 summarizes operating costs for a 600 MW regenerative operation (new and retrofit) with an 80% load factor. The case of retrofit with a 40% load factor is included to show the effect of this important variable on the system operating cost. The same relative behavior is observed with the operating costs as with capital costs -- temperature control by minimum air/fuel gives the lowest operating cost, while steam injection gives the highest. Steam injection is inferior because of the low overall efficiency inherent in the scheme at an air/fuel ratio of 20% of stoichiometric. The analysis points out that operating at the lowest possible air/fuel ratio, at which sulfur removal remains high and carbon deposition does not become excessive, gives the lowest operating and capital costs. All of the proposed temperature control schemes will, of course, approach the costs for

TABLE 2-9

REGENERATIVE OPERATION CAPITAL COSTS - 600 MW

EQUIPMENT X 10^{-3} \$	TEMPERATURE CONTROL METHOD				
	Heat Transfer Surface	Minimum Air/Fuel	Stack Gas Recycle	Steam Injection	Water Injection
Gasifier	1,020[a]	310	850	870	560
Regenerator[b]	135	135	138	144	140
Limestone handling	1,225	1,225	1,235	1,315	1,260
Fans	300	200	400	400	300
Burners	720	610	900	890	780
Particulate clean-up[c]	1,620	1,510	2,030	1,980	1,760
Boiler modifications	900	900	900	900	900
Equipment	5,920	4,890	6,453	6,499	5,700
Installation	2,960	2,445	3,227	3,250	2,850
Direct cost	8,880	7,335	9,680	9,749	8,550
Eng., cont., const.	2,660	2,200	2,910	2,930	2,570
Sulfur recovery	3,300	3,300	3,340	3,540	3,440
Total investment	14,840	12,835	15,930	16,219	14,560
Retrofit \$/kw	24.7	21.4	26.6	27.0	24.2
Retrofit \$/kw (without particulate control)	19.5	16.5	20.0	20.6	18.6
New \$/kw (without particulate control)	14.3	11.6	14.2	14.8	13.1

(a) Gasifier cost doubled to include heat transfer requirement.

(b) Regenerator temperature control by heat transfer surface doubles vessel cost.

(c) Particulate clean-up includes cyclones prior to burners and electrostatic precipitator final clean-up.

TABLE 2-10

OPERATING COSTS FOR REGENERATIVE OPERATION - 600 MW - 80% LOAD FACTOR

	TEMPERATURE CONTROL METHOD				
	Heat Transfer Surface	Minimum Air/Fuel	Stack Gas Recycle	Steam Injection	Water Injection
Capital Charges 10^3 \$/yr(a)	1620	1340	1770	1780	1560
Maintenance(a)	580	480	630	640	560
Overhead(a)	460	380	510	510	450
Limestone (2.5/ton)	278	278	280	296	284
Disposal (\$1.30/ton)	145	145	146	154	148
Sulfur recovery	1210	1210	1210	1270	1240
Fuel adder (R3/bbl)(b)	928	855	1010	2940	1580
	5221	4688	5556	7590	5822
Retrofit fuel adder,(c) ¢/10^6 Btu (80% load factor)	12.4	11.2	13.2	18.1	13.9
(40% load factor)	18.8	16.4	20.1	25.0	20.0
New design fuel adder, ¢/10^6 Btu (80% load factor)	10.7	9.5	11.1	16.0	12.0

(a) Includes particulate control.

(b) Based on overall efficiency given in Table 2-8.

(c) No sulfur credit. Credit would amount to about 1.5¢/10^6 Btu at \$15/ton of sulfur.

the minimum air/fuel ratio as the air/fuel ratio is reduced from the ratio of 20% of stoichiometric on which Tables 2-9 and 2-10 are based.

The three most important factors in determining the economics of the regenerative operation are (1) the carbon deposition rate, (2) the air/fuel ratio, and (3) the overall system efficiency. These three factors are interrelated, with the air/fuel ratio probably the single most important factor. The carbon deposition rate controls the cost of sulfur recovery and strongly affects the thermal efficiency of the system. The air/fuel ratio determines the equipment sizes and particulate clean-up costs, and also affects the thermal efficiency of the system. The overall efficiency controls the direct fuel adder cost and is itself strongly dependent on the carbon deposition rate, the air/fuel ratio, and the temperature control method.

Once-Through Operation Economics - New and Retrofit

Table 2-11 summarizes the capital cost breakdown for 600 MW once-through operation, new and retrofit. Temperature control by heat transfer surface, minimum air/fuel ratio, and stack gas recycle are considered, and the total capital investment in \$/kw is shown for retrofits with and without particulate control, and for a new boiler excluding the cost of particulate control. Again, operation with the minimum air/fuel ratio has the lowest capital cost, and the capital investment is lower for the once-through operation than for a regenerative operation.

Table 2-12 shows operating costs for the three temperature control schemes in ¢/10^6 Btu. Both 80% and 40% load factors are considered in the table, and the results for retrofits and new designs are shown. As expected, operating costs are lowest with temperature control by operation at the minimum air/fuel ratio, but operating costs for the once-through operation and the regenerative operation are extremely close. This indicates that the cost incurred by increasing the limestone feed rate by a factor of three, in going from regenerative to once-through operation, does not counteract the reduction in cost obtained by ceasing to recover sulfur.

With once-through operation, the air/fuel ratio controls the equipment sizes and the particulate clean-up costs and affects the system efficiency, as in the regenerative operation. As in the regenerative operation, the overall efficiency determines the direct fuel adder cost, but with once-through operation the rate of carbon deposition plays a much less important part in the system economics. This is because sulfur recovery is excluded, and the thermal efficiency is independent of the carbon deposition rate.

TABLE 2-11

ONCE-THROUGH OPERATION CAPITAL COSTS - 600 MW

	TEMPERATURE CONTROL METHOD		
Equipment x 10^{-3} \$	Heat Transfer Surface	Minimum Air/Fuel	Stack Gas Recycle
Gasifier	1,020	310	850
Sulfate generator	188	188	188
Limestone handling	1,320	1,320	1,330
Fans	300	300	400
Burners	720	610	900
Particulate clean-up	1,820	1,710	2,230
Boiler modifications	900	900	900
Equipment	6,268	5,338	6,798
Installation	3,134	2,669	3,399
Direct cost	9,402	8,007	10,197
Eng., cont., const.	2,820	2,400	3,060
Total investment	12,222	10,407	13,257
Retrofit, \$/kw	20.4	17.3	22.1
Retrofit, \$/kw (without particulate control)	15.2	12.4	15.5
New, \$/kw (without particulate control)	10.0	7.5	9.7

TABLE 2-12

OPERATING COSTS FOR ONCE-THROUGH OPERATION - 600 MW - 80% LOAD FACTOR

	TEMPERATURE CONTROL METHOD		
	Heat Transfer Surface	Minimum Air/Fuel	Stack Gas Recycle
Capital charges 10^3 \$/yr[a]	1710	1460	1855
Maintenance[a]	610	520	665
Overhead[a]	490	420	530
Limestone (\$2.50/ton)	835	835	840
Disposal (\$1.30/ton)	435	435	435
Fuel adder (\$3/bbl)[b]	830	740	928
	4910	4410	5253
Retrofit fuel adder ¢/10^6 Btu (80% load factor)	11.7	10.5	12.5
(40% load factor)	18.4	16.2	19.7
New design fuel adder ¢/10^6 Btu (80% load factor)	9.9	8.9	10.4

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[a] Includes particulate control.

[b] Based on overall efficiency given in Table 2-8.

TECHNICAL EVALUATION

The gasification/desulfurization concepts can be evaluated on the basis of the preliminary economic studies and technical feasibility investigations. The regenerative and once-through modes of operation are compared, and the merits and disadvantages of each mode are pointed out. The costs and pollution control potential of residual oil gasification/desulfurization is compared with the low-sulfur oil and stack gas cleaning alternatives. In addition, preliminary design conclusions are listed to summarize the results of the technical feasibility studies.

Comparison Between Regenerative and Once-Through Operations

The relative merits of the two modes of operation are summarized in the following matrix. Preliminary investigations indicate that, based on all of the points in this matrix, the once-through operations may be somewhat more attractive to a utility customer than the regenerative operation. Further research into the areas of sulfur recovery and system dynamics may alter this outlook.

COMPARISON BETWEEN MODES OF OPERATION

Economics

- Once-through operation requires lowest capital investment
- Operating costs are nearly identical for once-through and regenerative operations, mainly because of the high cost associated with sulfur recovery
- Sulfur credit does not greatly reduce the operating cost of a regenerative operation

Space Requirements

- Both modes have the same potential equipment sizes, though the regenerative operation requires a large space for sulfur recovery, making the once-through more compact overall
- Once - through may require deeper gasifier bed depth and larger limestone handling equipment

Performance

- Both modes have same potential for sulfur removal
- Once-through may be slightly more efficient than regenerative
- Once-through has higher power requirements
- Once-through thermal efficiency is independent of carbon deposition rate
- Regenerative operation minimizes stone disposal rate
- Once-through at present has fewer technical problems

Utility Reaction Factors

- Regenerative mode puts utility in the chemical business
- Once-through is the simplest mode of operation, but may increase solid waste problem unless high stone utilization can be achieved

Comparison with Other SO_2 Control Methods

Table 2-13 compares atmospheric pressure oil gasification/desulfurization with the alternative schemes of low-sulfur oil and stack gas cleaning. Capital costs and operating costs are compared for new and retrofit systems. Oil gasification/desulfurization compares favorably with low-sulfur oil and stack gas cleaning, based on the preliminary cost estimates presented in Tables 2-9 through 2-12. A reduction of about 40% in the capital costs involved in stack gas cleaning is estimated for new and retrofit gasification/desulfurization systems. Operating costs appear to be about the same for stack gas cleaning and gasification/desulfurization, while a reduction of about 20% in the operating cost for the low-sulfur fuel case by use of the gasification/desulfurization system is estimated. These conclusions are based on the desulfurization of high-sulfur residual oil (3 wt % sulfur) and may be altered when a lower sulfur oil is considered (1 to 1.5 wt % sulfur). Environmental factors are also compared in Table 2-13. The low-sulfur oil alternative is advantageous in that capital costs are limited to possible boiler modifications necessary when changing from gas or coal to low-sulfur oil. On the other hand, operating costs are higher than for stack gas cleaning or oil gasification/desulfurization. Capital costs are extremely high with stack gas cleaning, especially in the retrofit case, while operating costs are very near those estimated for oil gasification/desulfurization.

TABLE 2-13

ASSESSMENT OF FLUIDIZED BED OIL GASIFICATION-DESULFURIZATION(a)

	LOW-SULFUR OIL	STACK GAS CLEANING	OIL GASIFICATION	
			Regenerative Operation	Once-Through Operation
COST				
Capital, $/kw				
New	---	25	12-15(c)	8-10(c)
Retrofit	(b)	40-75	22-27(c)	18-22(c)
Fuel adder, ¢/10^6 Btu				
New	17-23	11	9.5-16.0(c)	9-10.5(c)
Retrofit	17-23(b)	14-20	11-18	10.5-12.5(c)
EFFICIENCY	---	0.95-0.98	0.89-0.96(d)	0.96-0.97(d)
ENVIRONMENTAL FACTORS				
SO_2, ppm (lb/10^6 Btu)	---	100-110 (0.45)	80-100 (0.35)	80-100 (0.35)
NO_x, ppm (lb NO_2/10^6 Btu)	---	400-700 (0.8)	100-150 (0.16)	100-150 (0.16)
Particulates, gr/SCF (lb/10^6 Btu)	0.03 (0.06)	0.03 (0.05)	0.01-0.10(e) (0.02-0.2)	0.01-0.10(e) (0.02-0.2)
Solid waste, ft^3/MW-day	---	25	15	45
S removal				
Stone	NA		Limestone	Limestone
Ca/S	NA		~ 15(f)	3.0
Make-up Ca/S	NA		1.0	NA

(a) Basis: 3% sulfur, 90% sulfur removal, 600 MW capacity.

(b) Equipment modifications are required when converting from gas or coal to low-sulfur oil.

(c) See Tables 2-9 through 2-12 for details.

(d) Overall efficiency is dependent on mode of temperature control; see Table 2-8.

(e) 0.01 figure based on installing electrostatic precipitator (ESP).
0.1 figure based on installing high-efficiency cyclone before burners and no ESP.

(f) Ca/S ratio dependent on regenerator temperature control scheme.

Preliminary Design Conclusions

Preliminary design conclusions listed in the following matrix are based on the technical investigations conducted. Though the analysis has raised many technical questions, the preliminary investigation coupled with the experimental work of Esso (England) points out the general feasibility of oil gasification/desulfurization as a retrofit SO_2 control system for utility boilers.

PRELIMINARY DESIGN CONCLUSIONS

TECHNICAL PROBLEM AREAS

Sulfur recovery

Degradation of boiler performance and boiler modifications

Calcium sulfate generation

Temperature control

Lime circulation

CURRENT OUTLOOK

Overall, once-through operation appears superior to regenerative if high stone utilization can be achieved, though this is not definite. (See page 526).

Gasifier temperature control by means of the minimum air/fuel ratio appears superior to other methods, if it can be achieved in practice. Temperature control with heat transfer surface also may be attractive if high air/fuel ratios ($\geq$ 20%) must be used in operation.

The regenerator temperature is best controlled by stone circulation if carbon deposition is small. If carbon deposition is large, then H.T.S. is the most efficient method.

Sulfate generator temperature is most easily controlled by circulating excess air through the generator.

The situation concerning the internal and external design concept is not clear (page 507) with the number of variable factors involved. It seems reasonable to strive to minimize the expense and time involved in boiler modifications.

The greatest present market for oil gasification/desulfurization appears to be in retrofits. The problems involved with retrofits should be examined in more detail.

Oil gasification/desulfurization concepts appear highly competitive with other retrofit or new SO_2 control methods based on preliminary designs (Table 2-13) and should be investigated in greater detail. (See recommendations).

RECOMMENDATIONS

The performance and cost for a utility boiler add-on system are competitive with stack gas cleaning and with the purchase of clean oil. Based on this technical assessment, we recommend that

- The market for oil for power generation beyond 1975-1980 continue to be assessed in order to evaluate the market potential for atmospheric and pressurized oil gasification.
- Utilities be contacted and a utility partner selected for a demonstration installation of an add-on oil gasification unit.
- A preliminary design of the demonstration plant be prepared.
- Research and development work be continued in key areas -- e.g., carbon deposition, temperature control, load range, sulfur recovery, pressurized operation.

3. FLUIDIZED BED COMBUSTION DEVELOPMENT PLANTS

DEVELOPMENT PLANT ALTERNATIVES

Three advanced fluidized bed power generation systems and an industrial fluidized bed steam generation system have been studied; these system concepts are summarized in the following matrix. Experimental development and systems evaluation have been carried out for all three.

FLUIDIZED BED FUEL PROCESSES FOR STEAM-POWER GENERATION

	FLUIDIZED BED COMBUSTION	FLUIDIZED BED GASIFICATION
Atmospheric Pressure	•Replacement for conventional utility boiler with SO_2 stack gas cleaning system and NO_x control burning high-sulfur coal or oil. •Industrial boiler burning high-sulfur coal.	•Substitute for SO_2 stack gas cleaning system and NO_x control on utility plant burning high-sulfur oil.[a]
10-30 Atm Pressure	•Pressurized fluid bed boiler operated with combined gas and steam turbine-generators burning high-sulfur coal or oil.	

[a] New plant or retrofit on Existing plant.

All of these systems offer potential benefits, and each system might be developed independently. However, the selection of the system(s) to be developed requires the assessment of both technical and nontechnical factors. The potential for pollution abatement, ability to use readily available fuels, promise for economical steam/power generation, present state of development, technical problem areas, advance design potential,

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and projected development time and cost have been assessed for each system and compared with conventional fossil fuel-fired steam/power generating plants utilizing present pollution abatement technology. Other factors, such as the level of funding, alternative technological advancements, future energy usage, future fuel prices, and future regulations, e.g., emission standards and resource allocations, must also be considered.

In order to recommend a fluidized bed fuel processing development plant, the respective system evaluations and experimental development and market projections were studied with respect to application -- utility/industrial, new/retrofit, and technology -- combustion/gasification and pressurized/atmospheric pressure. Recommendations are based on the following assumptions:

- Available funds will be limited
- Alternative technological advancements will not displace the use of fossil fuels as a major electric utility fuel in the next 50 to 100 years
- Utility concepts should be demonstrated within the next 10 to 15 years
- Future fuel combustion emissions standards for SO_2, NO_x, and particulates will require emissions less than or equal to the presently proposed national standards.

On the basis of the evaluations and the assumptions, it is recommended that

- Pressurized fluidized bed combustion be developed for new electric utility plants
- A pressurized fluidized bed combustion development plant be constructed and operated
- A retrofit atmospheric pressure oil gasification demonstration plant be constructed and operated by a utility.

PRESSURIZED FLUIDIZED BED COMBUSTION BOILER DEVELOPMENT PLANT

Objectives

The pressurized fluid bed combustion boiler program at BCURA demonstrated the feasibility of pressurized fluid bed combustion in an 8 ft^2 combustor. Specifically, it demonstrated:

- SO_2 emissions of ≤ 0.7 lb/10^6 Btu
- NO_x emissions of 0.07-0.2 lb NO_2/10^6 Btu
- Coal feeding at 3-1/2 and 5 atm
- Continuous operation - runs up to 350 hours terminated by choice
- No erosion or corrosion of gas turbine blade test passages after 200-hour tests
- Particulate removal at ∿ 1500°F and 5 atm by cyclones proved adequate for turbine blade tests
- Boiler tube materials
- Bed operated with horizontal tube bundle.

The Westinghouse-Foster Wheeler boiler design operating conditions require that the bed be operated at higher gas velocity, higher bed temperature, higher pressure, and with a deeper bed than the BCURA unit. Each bed area for a four--module 300 MW plant is 35 ft^2. The development plant is required to investigate the performance of the proposed boiler plant equipment system design at the proposed operating conditions for the following features:

- SO_2 and NO_x emissions
- Operation of deep beds (10 to 20 ft) with horizontal tube bundles and headers
- Particulate control equipment
- Multiple coal feeding and sorbent distribution in a large bed (∿ 35 ft^2)
- Materials and component life, including boiler tubing and gas turbine blades

- Sorbent regeneration processes and compatibility of sulfur-rich gas with sulfur recovery processes
- Operational techniques - start-up, shut-down, load follow
- Long-term operability
- Sorbent circulation.

Other objectives which should be considered when planning the facility include:

- Need to test sulfur recovery system
- Feasibility of studying advanced concepts - higher steam temperature and pressure, higher gas turbine temperatures, circulating beds, deeper beds (30 ft), higher pressures
- Feasibility of studying components of a pressurized gasification system
- Feasibility of expanding to multiple bed operation.

The development plant should be designed so that sufficient information can be obtained to design, build, and operate a demonstration plant of 150 to 300 MW.

Plant Concept

For the plant to achieve these objectives, the following design features must be provided:

- Capability of operating over the following range of projected operating conditions: pressure up to 20 atm; bed temperature, 1300°F to 2000°F; gas velocity, 6 to 15 fps; bed depth, 10 to 30 ft
- Bed area large enough to (1) test multiple-point coal feeding, (2) avoid excessive aspect ratios (L/D) with deep bed operation, (3) test proposed heat transfer surface configurations
- Gas flow sufficient to obtain turbine blade test data on a rotating turbine - minimum practical size 3 to 5 MW
- Capacity and configuration of sufficient size that further development work will not be required -- e.g., primary bed capacity should be designed to permit carbon burn-up cell design to be tested; minimum scale-up to demonstration plant
- Sufficient capacity to permit study of regeneration processes on a scale large enough to demonstrate practicality.

These design characteristics can be achieved by constructing a fluidized bed carbon burn-up cell unit equivalent to the bottom bed of a module in the four-module, 300 MW boiler plant design. The primary bed area is 35 ft^2, which would permit a realistic test of fuel feeding. The regenerator vessels for such a unit would be approximately two feet in diameter, based on the process design for the 635 MW plant. Regenerator vessels of this size are expected to provide the necessary design data for a larger plant. Since equivalent capacity of the unit operating at 10 atm, 1750°F, 10 fps, and 10% excess air would be $\sim$ 30 MW, a gas turbine of about 5 MW could be used to test turbine blades. The bed would be a modular unit of the 300 MW plant design, so the scale-up to a demonstration plant would only require going to a multiple bed operation. The primary bed carbon burn-up cell unit would be placed in a shell large enough to provide easy access to achieve design modifications and to study alternative concepts. Design flexibility would permit modifications in the heat transfer surface. The combustor could operate as either an evaporator, superheater, or reheater, but could not perform all functions together. The plant requires ∿ 10 tons of coal per hour and would be designed to handle up to ∿ 500,000 lb/hr of water or steam.

The location of the plant is an important consideration, since large quantities of coal and water are required. Several advantages could be achieved by locating the development plant on an existing power plant site:

- Availability of coal handling, water preparation, and solids disposal facilities
- Existing plant would dispose of steam
- Superheat and reheat steam generation can be studied by using bleed stream from existing plant
- Utility partnership would minimize site development and development time, and provide plant utilities and maintenance facilities.

A partnership with a utility is therefore recommended.

Conceptual Design

Flow Diagram and Material Balance

Figure 3-1 is a flow diagram of the pressurized fluid bed combustion boiler development plant. The plant is adjacent to a large power plant to which it has access to provide interfaces with the coal, water, steam, waste stone, and utilities. The facility is designed to operate over the range of operating conditions presented on page 537 .

The material balance and operating conditions for the development plant at the projected 100% load design conditions are presented in Table 3-1. The material balance is based on the coal and dolomite compositions presented in the specification list on page 542 and Table 3-2. The boiler is designed to handle up to 500,000 lb/hr of steam flow. The unit would function as either a pre-evaporator, superheater, or reheater. Water evaporation in the walls could also be studied.

Equipment

Boiler. The boiler is a single fluidized bed with a carbon burn-up cell. The design is based on the one proposed by Westinghouse and Foster Wheeler. The water-walled, fluidized bed boiler would be placed in a pressure shell larger than required for commercial operation. A vessel 15 to 20 feet in diameter and 50 to 70 feet high is projected to permit easy access, flexibility for modifications, and operation over a wide range of operating conditions.

Coal and Dolomite Handling System. Coal-receiving equipment and storage is assumed to be available from the power plant. Equipment specified for the development plant includes a dolomite receiving hopper, dolomite storage silo, dolomite crusher, coal dryer, and coal crusher. The moisture content of the solids must be maintained at less than 3% to

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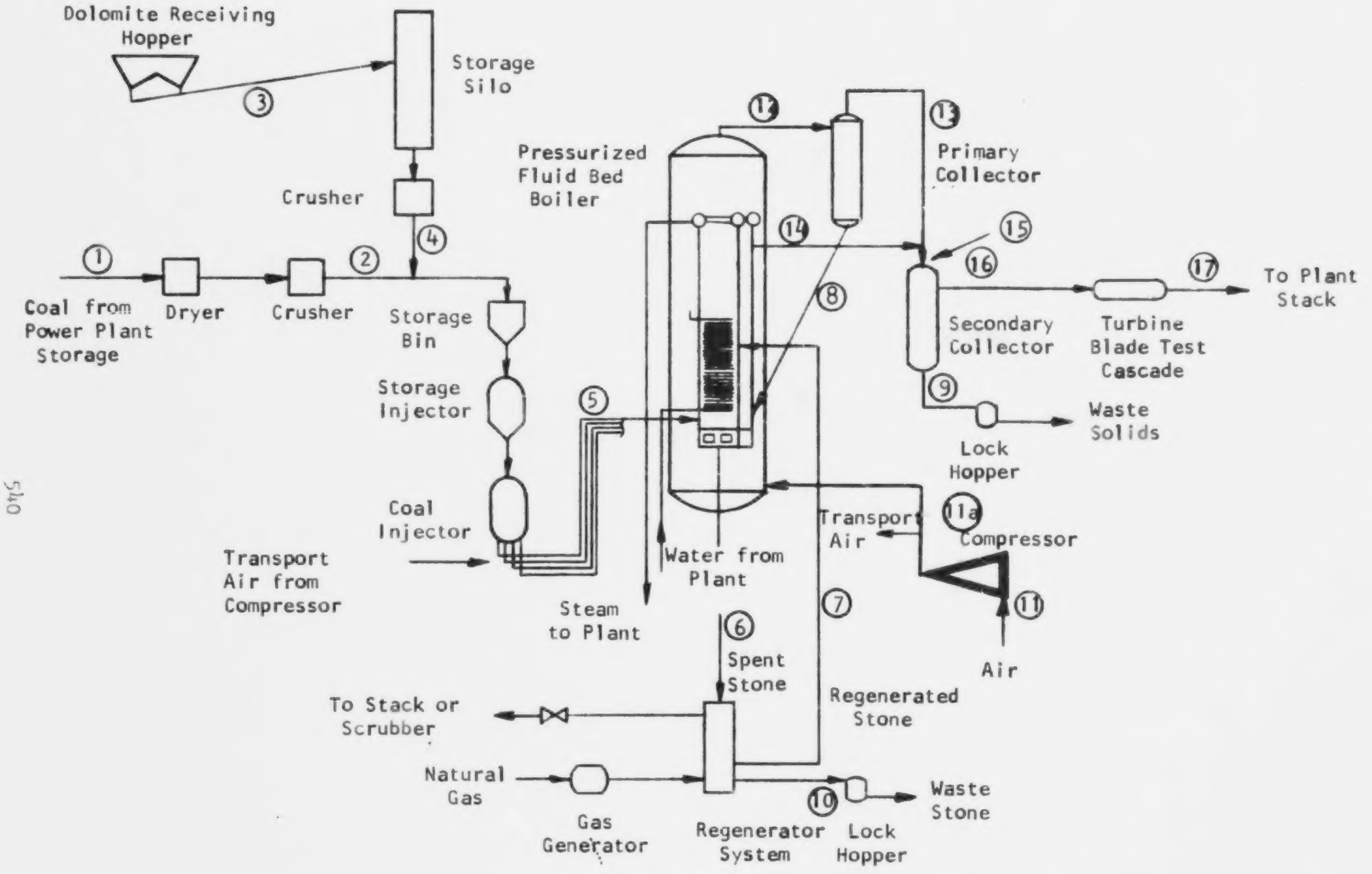


Location: Existing Power Plant

Fig. 3-1—Flow diagram for development plant facility

TABLE 3-1

MATERIAL BALANCE - DEVELOPMENT PLANT

STREAM	DESCRIPTION	TEMPERATURE °F	PRESSURE psia	FLOW RATE lb/hr	COMPOSITION wt %	PARTICLE SIZE
1	Coal from power plant	ambient	15	66,000[a]	Page	1-1/2" x 0
2	Sized coal	100	15	66,000[a]	Page	1/4" x 0
3	Dolomite	ambient	15	--	Table 3-2	1-1/2" x 0
4	Dolomite	ambient	15	6,000[a]	Table 3-2	1/4" x 0
5	Coal/dolomite to boiler	~ 500	200	24,000	Streams 2 & 4	1/4" x 0
6	Spent dolomite	1750	150	20,000	CaO-MgO 75 $CaSO_4$-MgO 25	-1/4"
7	Regenerated dolomite	1100	150	24,000	CaO-MgO 3 $CaCO_3$-MgO 94 $CaSO_4$-MgO 3	-1/4"
8	CBC feed	1600	142	2,900	Ash 60 Carbon 30 Dolomite 10	-1/32"
9	Waste solids	1600	139	2,000	Ash 80 Carbon 6 Stone 14	-1/32"
10	Waste solids	1100	180	2,300	same as 7	-1/4"

				FLOW RATE lb/sec	FLOW RATE ACFM	COMPOSITION mole %	SOLIDS LEADING gr/SCF
11	Air	ambient	15	75	56,600	N_2 77.5 O_2 20.9 H_2O 2.1	nil
11a	Air	600	150	75	11,100	same as 11	nil
12	Flue gas from FBB	1700	145	73	21,800	CO_2 15 N_2 74 CO 0.2 H_2O 9 O_2 2 NO_x ~200 ppm SO_2 ~150 ppm	7-15
13	Flue gas from primary collector	1650	142	73	21,800	same as 12	~1
14	Flue gas from CBC	1750	142	8	2,380	same as 12	4-7
15	Flue gas to secondary collector	~1650	142	81	24,200	same as 12	5-8
16	Flue gas to test passage	1600	139	81	24,200	same as 12	<0.15
17	Flue gas to stack			81		same as 12	<0.15

(a) Assumes crushers and dryer operate one shift per day.

SPECIFICATION LIST
OHIO PITTSBURGH NO. 5 SEAM COAL
(Source of data: USBM, Pittsburgh, Pa.)

SAMPLE: Run of mine - as received

PROXIMATE ANALYSIS (wt %):

Moisture	3.3
Volatile matter	39.5
Fixed carbon	48.7
Ash	8.5
	100.0

ULTIMATE ANALYSIS (wt %):
(includes moisture)

C	71.2	
H	5.4	
O	9.3	
N	1.3	
S	4.3	(∿60% organic; ∿40% pyritic)
Ash	8.5	
	100.0	

GROSS HEATING VALUE: 13000 Btu/lb

NET HEATING VALUE: 12500 Btu/lb

ASH ANALYSIS (wt %):

SiO_2	45.3
Al_2O_3	21.2
Fe_2O_3	27.3
TiO_2	1.0
P_2O_5	0.11
CaO	1.9
MgO	0.6
Na_2O	0.2
K_2O	1.8
SO_3	0.7
	100.1

FUSIBILITY OF ASH:

Initial deformation temperature	-- 2080°F
Softening temperature	-- 2230°F
Fluid temperature	-- 2420°F

PARTICLE DENSITY:
Coal -- ∿1.4 gm/cc
Ash -- ∿2.8 gm/cc

GRINDABILITY (Hardgrove): 50-60

FREE SWELLING INDEX: 5-5.5

TABLE 3-2

SO_2 SORBENT ANALYSIS USED FOR PRELIMINARY FLUIDIZED BED BOILER ANALYSES

Component	wt % As Received[a] Dolomite 1337
SiO_2	0.78
Al_2O_3	0.15
Fe_2O_3	0.25
MgO	45.0
CaO	53.0
TiO_2	0.02
SrO	< 0.03
Na_2O	< 0.02
K_2O	< 0.1
MnO_2	< 0.03

[a] R. W. Coutant, J. S. McNulty, R. E. Barrett, J. J. Carson, R. Fischer, and E. H. Lougher "Investigation of the Reactivity of Limestone and Dolomite for Capturing SO_2 from Flue Gas," August 1968 (Battelle Memorial Institute).

prevent agglomeration in the pressurized coal injection equipment. A fluidized bed dryer is therefore included in the coal handling system to reduce the moisture picked up during handling and storage to ≤ 3%. Since the dolomite is received by covered railroad car and is assumed to contain less than 3% moisture, no drying of the dolomite is required. Reversible hammermill crushers are used to reduce the coal and dolomite from 1-1/2 in x 0 to 1/4 in x 0.

Coal Feed System. The Petrocarb pressurized coal feed system is used to feed the coal and make-up dolomite to the fluid bed boiler. The system includes a storage bin, storage injector, and coal injector which supplies coal and dolomite to the bed at four locations. More feed points can be installed if required.

Particulate Removal. The primary particle collector is a cyclone designed to achieve 95% particle collection efficiency by weight. The secondary collector is a tornado unit designed to achieve 97% to 99% collection efficiency.

Turbine Blade Test System. Initial tests will be carried out with a turbine blade test cascade. The second phase of operation will include a rotating gas turbine test unit.

Air Compressor. An air compressor with motor drive, independent of the gas turbine test system, has been selected for the development plant.

Regeneration System. The present conceptual design employs the two-step process specified in the pressurized fluid bed boiler power plant design. However, additional work on regeneration is recommended prior to construction of the regeneration system.

Auxiliary equipment cost estimates are based on the two-step process. Equipment includes a reducer vessel, H_2S generator vessel, solids handling system, gas generator for producing a reducing gas, and a source of CO_2/H_2O gas. The reducer vessel and H_2S generator vessel would each be approximately two feet in diameter. The reducing gas requirements would be approximately 2500 SCFM of gas, composition 16% CO, 8% H_2, 7% CO_2, 8% H_2O, 59% N_2. This gas would be supplied by a gas generator burning natural gas. The CO_2/H_2O gas requirement would be approximately 2000 SCFM of gas, composition 36% CO_2 and 64% H_2O. This gas is supplied by purchasing merchant CO_2 and blending it with steam. Absorption of CO_2 from the power plant stack gases should also be considered if this regeneration process is selected for further development.

Space Requirement

The development plant will require approximately 90 ft x 55 ft, or ∿ 5000 ft^2. This does not include space for coal storage, stack, or waste solid storage.

Power Requirement

The estimated power requirement for pumps, compressors, and crushers is 23,000 hp.

Experimental Program

The development plant program is broken down into three phases of operation:

Phase I. Operation of fluid bed boiler with turbine blade test cascade and without sorbent regeneration process

Phase II. Operation of integrated system, including gas turbine and regeneration process

Phase III. Operation to evaluate alternative concepts, e.g., regeneration processes, bed design, steam conditions.

The first phase would provide information to assess the proposed design concept—boiler and carbon burn-up cell. Design and operating characteristics studied during Phase I operation include:

- Operation of deep beds with proposed heat transfer surface design
- Temperature gradients, particle circulation, particle carry-over, heat transfer coefficients, combustion efficiency, SO_2 removal, and NO_x minimization at the projected operating bed height, temperature, pressure, particle size, and excess air

- Coal feeding system
- Particulate control equipment
- Grade efficiencies, pressure drop, reliability
- Gas turbine blade erosion/corrosion/deposition
- Boiler tube material corrosion/erosion/deposition
- Boiler start-up, shut-down, and load follow.

The second-phase operation would incorporate design modifications recommended from the analysis of Phase I data, and the parallel research and development efforts include a regeneration process and a small gas turbine unit. A sulfur recovery process should also be considered if development effort is required. Design and operating characteristics studied during Phase II include:

- Regeneration process
- Operation of the integrated system
- Solids circulation, sorbent stability and activity, and feasibility of sulfur recovery
- Gas turbine blades and boiler tubes
- Integrated system operation—start-up, shut-down, load follow.

Phase III would assess alternative concepts. Development of the third-phase program will rely on the results of the analysis of data from Phases I and II, results of pilot and bench-scale experimental programs, and recommendations from systems evaluation and advanced concept studies. This phase might include the evaluation of higher gas turbine temperatures, higher steam temperatures and pressures, alternative regeneration/sulfur recovery processes, and circulating bed boiler concepts.

Cost Estimate

The estimate for the development plant equipment, installed out-of-doors adjacent to a large central station power plant, is $7,000,000. A cost breakdown with the sources of the respective costs is presented in Table 3-3.

TABLE 3-3

DEVELOPMENT PLANT COST ESTIMATE

EQUIPMENT	DESCRIPTION	EQUIPMENT COST ESTIMATE	ESTIMATE SOURCE
Coal and dolomite handling system	Hopper(s) Storage silo(s) Crusher(s) Dryer Conveyors	300,000	Westinghouse(a) Heyland Patterson Inc. Koppers
Coal Feed System		200,000	Petrocarb, Inc.
Storage bin	10' x 10' x 35' ht		
Storage injector	6' ϕ, 15' ht		
Coal injector	8' ϕ, 15' ht		
Pressurized FBB			
Boiler internals	Heat transfer surface in one bed of 300 MW modular design	350,000	Foster Wheeler Corp.
Shell	15-20' ϕ, 50-70' ht	100,000	Westinghouse
Particulate removal		300,000	
Primary	Cyclone(s)		Ducon - cyclone
Secondary	Tornado unit		Aerodyne - tornado
Turbine blade test cascade		50,000	Westinghouse
Air compressor and motor		800,000	Westinghouse Allis Chalmers
Regeneration system	Reducer vessel H_2S generator vessel Gas generator for reducing gas CO_2 supply	200,000	Westinghouse
Subtotal		2,300,000	
Engineering, piping, erection, installation, instrumentation, controls		4,700,000	Westinghouse(a)
TOTAL		7,000,000	

(a) Architecture and engineering firms were consulted.

Development Plant Schedule

A development plant schedule is presented in Figure 3-2. The projected time required to assess the combustor design, develop a regenerator process, and obtain data on turbine blade performance is four years. The plant could then be used to study alternative boiler and regeneration system concepts.

The schedule includes time for selecting a site and design contractor. A utility site is preferred. The design is broken down into a preliminary design phase, followed by the detail design. Construction is estimated to take one year. The plant would be operated for two years to obtain the necessary information for the design of a demonstration plant. The operation would be divided into two phases: the first would comprise operation of the boiler, including the carbon burn-up cell, and a turbine blade test cascade; the second would study operation of a regeneration process and the integrated boiler-regeneration system. A gas turbine would also be installed to obtain data on a rotating unit.

A third phase of operation is recommended to evaluate alternative concepts for the boiler, regeneration process, and sulfur recovery. This phase is not considered necessary for design of a demonstration plant, however.

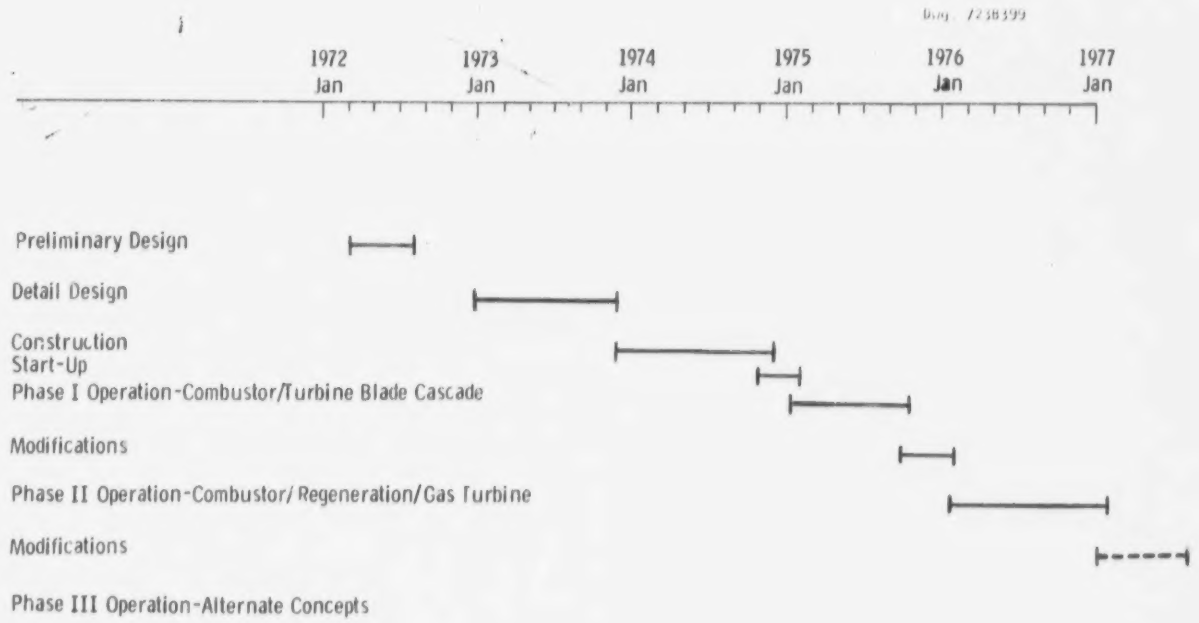


Fig. 3-2—10-30 MW pressurized fluid bed boiler development plant schedule

ATMOSPHERIC PRESSURE FLUIDIZED BED OIL GASIFICATION-COMBUSTION DEMONSTRATION PLANT

The preliminary design of the gasification/desulfurization retrofit demonstration plant is based on the conceptual designs and the material and energy balances presented in Section 2. Further experimental information for a preliminary design will be provided by the operation of the Esso (England) continuous pilot plant unit. The specific design of the utility boiler to be retrofit and the preliminary design will determine the retrofit concept and operational concepts to be applied to the demonstration plant. The cost of the development plant and the experimental program to be followed will depend largely on the specific design concepts applied (internal or external design, once-through or regenerative operation, temperature control schemes) and the characteristics of the utility boiler (size, fuel normally utilized, location, specifics of boiler design). Boiler capacities in the 50 to 200 MW range should be considered for the demonstration plant.

Objectives

The objective of the atmospheric pressure fluidized bed oil gasification-combustion demonstration plant is to demonstrate the technical and economic feasibility of the oil gasification/desulfurization concept in a utility boiler. The preliminary and detailed design phases will provide information about the costs and problems involved in boiler modification for the specific boiler from which general modification information can be implied. The experimental information gained from the operation of the plant will indicate the performance of the system (sulfur removal, NO_x, particulates, boiler performance, operating cost, etc.) and will permit the design of commercial retrofit units for boilers with characteristics other than those of the development boiler.

Plant Concept

The plant concept applied will depend on the specific boiler to be retrofit, the outlook of the utility partner, and the findings of the preliminary design phase. At present, most of the concepts proposed in Section 2 are possibilities for the demonstration plant operation. The alternative of retrofitting with the internal or the external design concept may be quickly decided once a utility partner is located, while the temperature control schemes to be applied to the demonstration plant may be chosen on the basis of the results from the Esso (England) pilot plant unit. The mode of operation (regenerative or once-through) to be applied may be a more difficult choice. The two alternatives will be evaluated with the utility and during the preliminary design phase.

Conceptual Design

Both regenerative and once-through operations are possibilities for the demonstration plant. The factors involved in the two modes of operation with respect to the demonstration plant will differ; and the equipment involved in the system, the specific experimental program carried out, the development plant cost, and the estimated time schedule for the development plant program will depend on which mode of operation is selected.

Flow Diagrams

Flow diagrams for regenerative and once-through operations were shown in Figure 3-1. The components in the flow diagrams are those discussed in the conceptual designs of Section 2. Conceptual plant layouts, based on these flow diagrams, are shown in Figures 3-3 through 3-6 for a 200 MW regenerative operation and a 200 MW once-through operation. The figures are drawn roughly to scale, with each gasification system consisting of two modules, though, a demonstration plant would use only one such unit. The figures show the essential vessels, fans, and streams, and the relative configuration of these components in the regenerative and once-through operations with the gasification/desulfurization system located external to the boiler.

Equipment

Equipment and boiler modifications are listed in the next matrix. Equipment normally present in an existing coal-fired boiler and an existing oil-fired boiler is indicated to show how capital investment might differ in the two cases. Boiler modifications required only with the internal design concept or the external design concept are also noted. Finally, the equipment needed only in the regenerative operation or only in the once-through operation is differentiated. Equipment present in both modes of operation, such as fans and lime transport lines, may differ in each operation, while some equipment, such as burners, will be identical in the two modes.

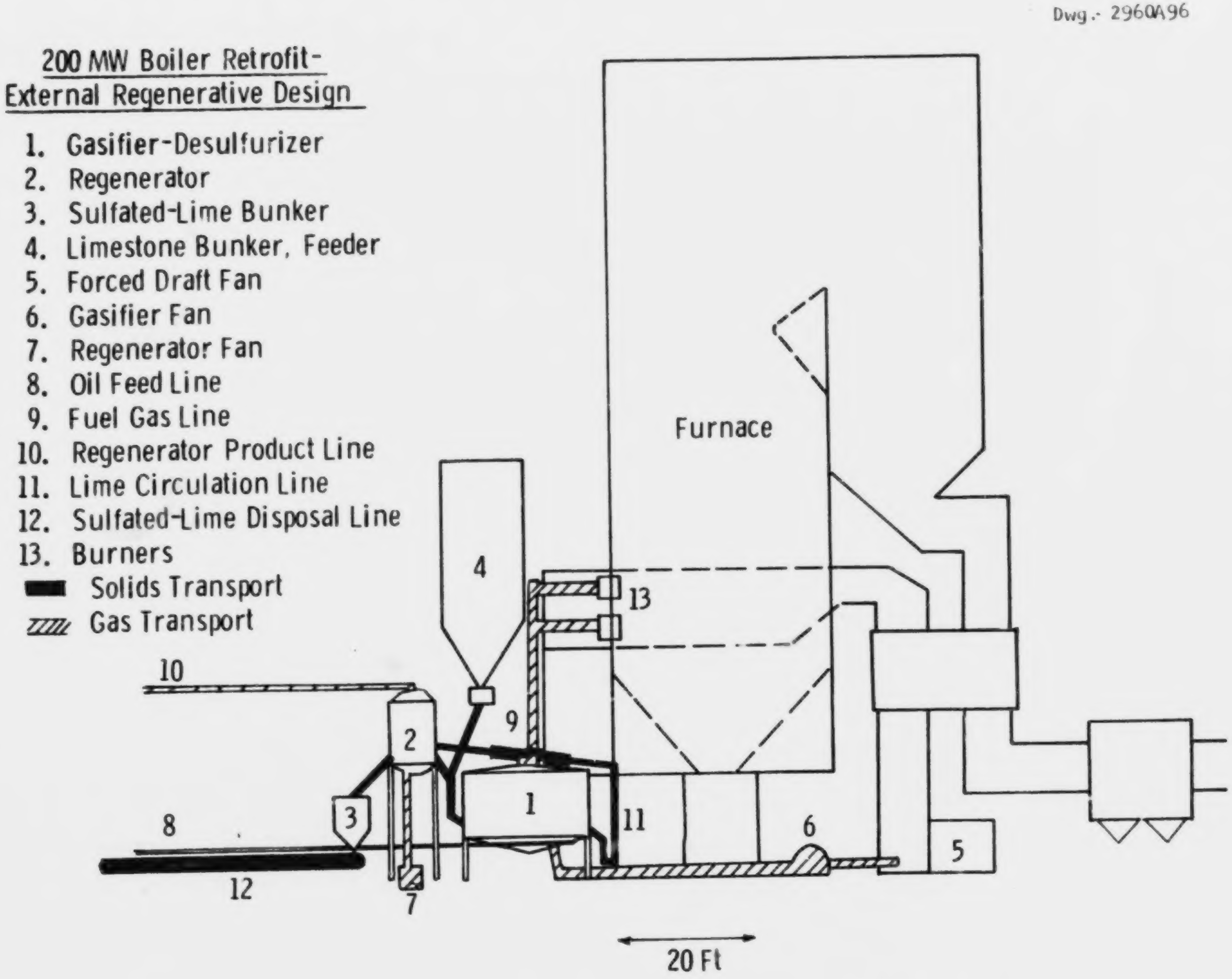


Fig. 3-3—Elevation of regenerative design-200 MW retrofit

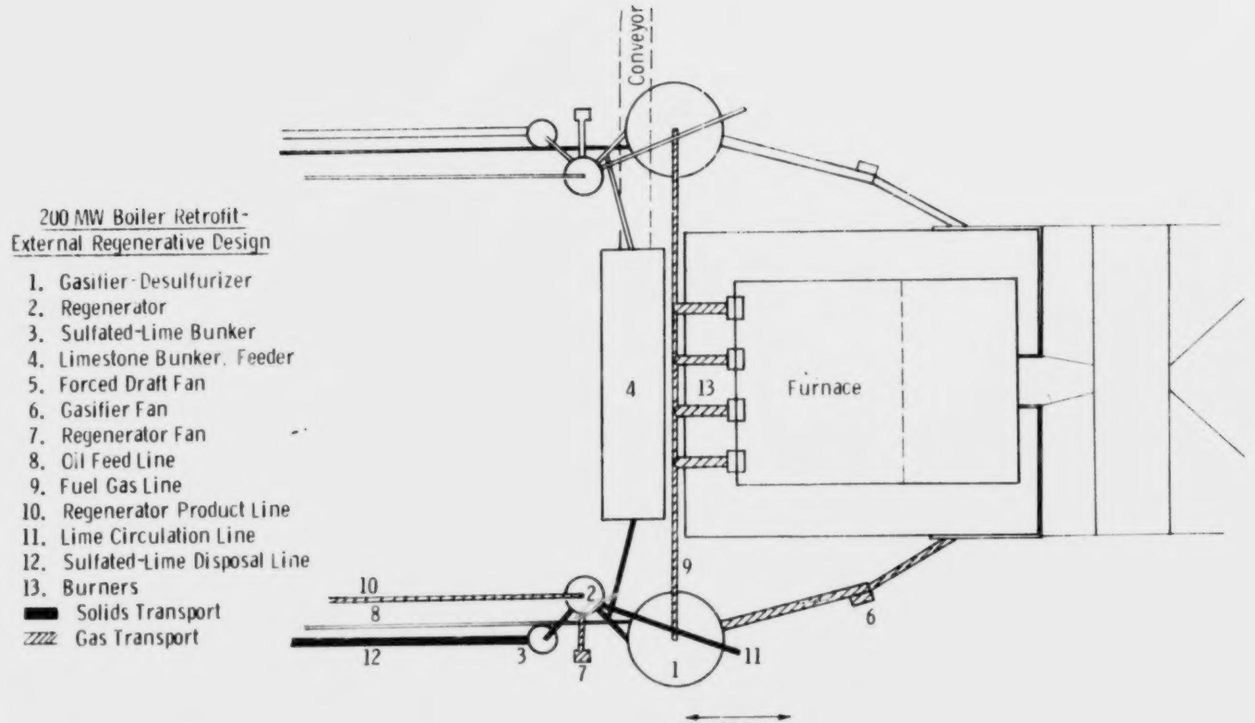


Fig. 3-4—Plan of regenerative design - 200 MW retrofit

Dwg. 2960A97

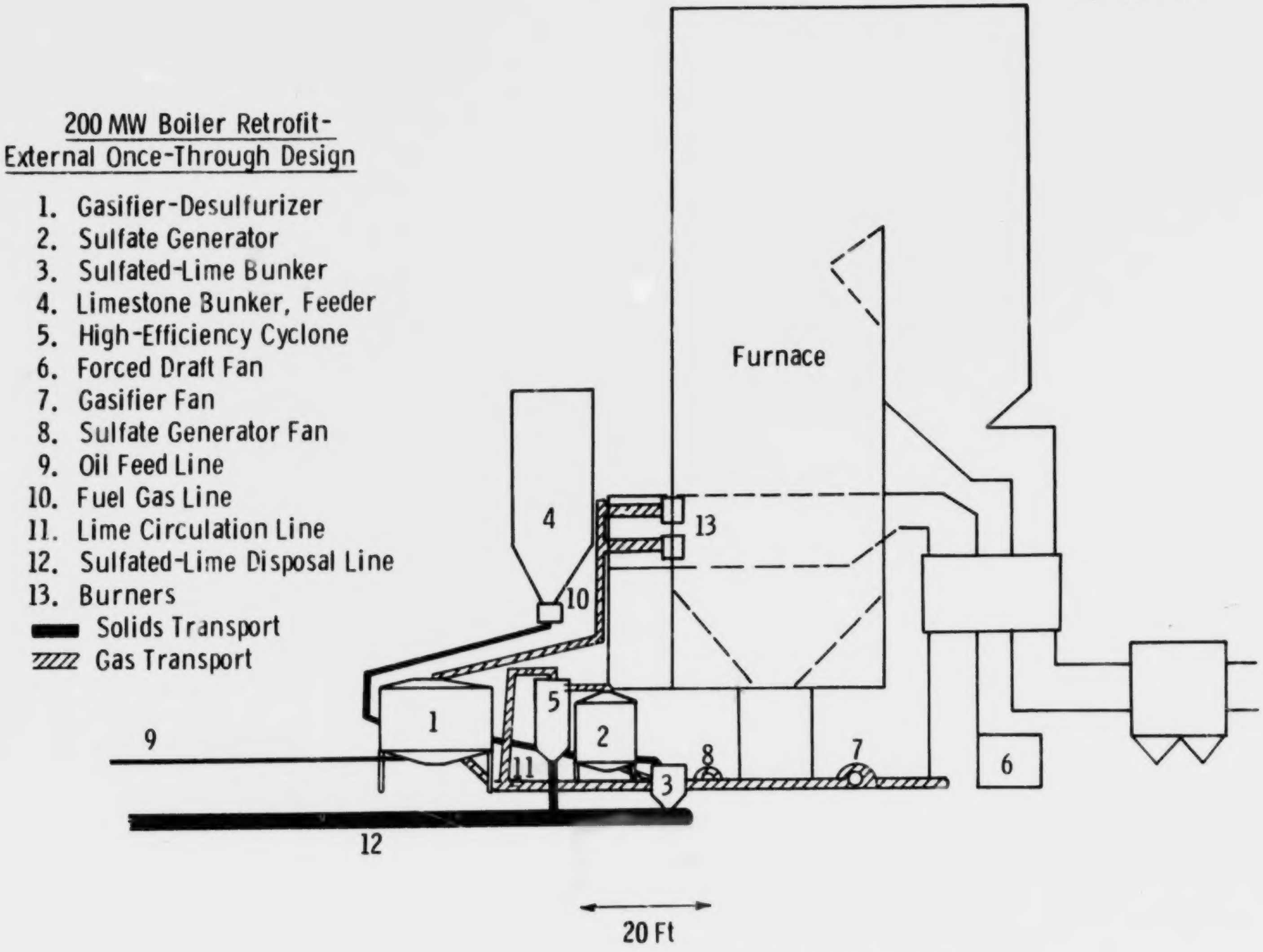


Fig. 3-5—Elevation of once-through design-200 MW retrofit

Dwg. 2960A98

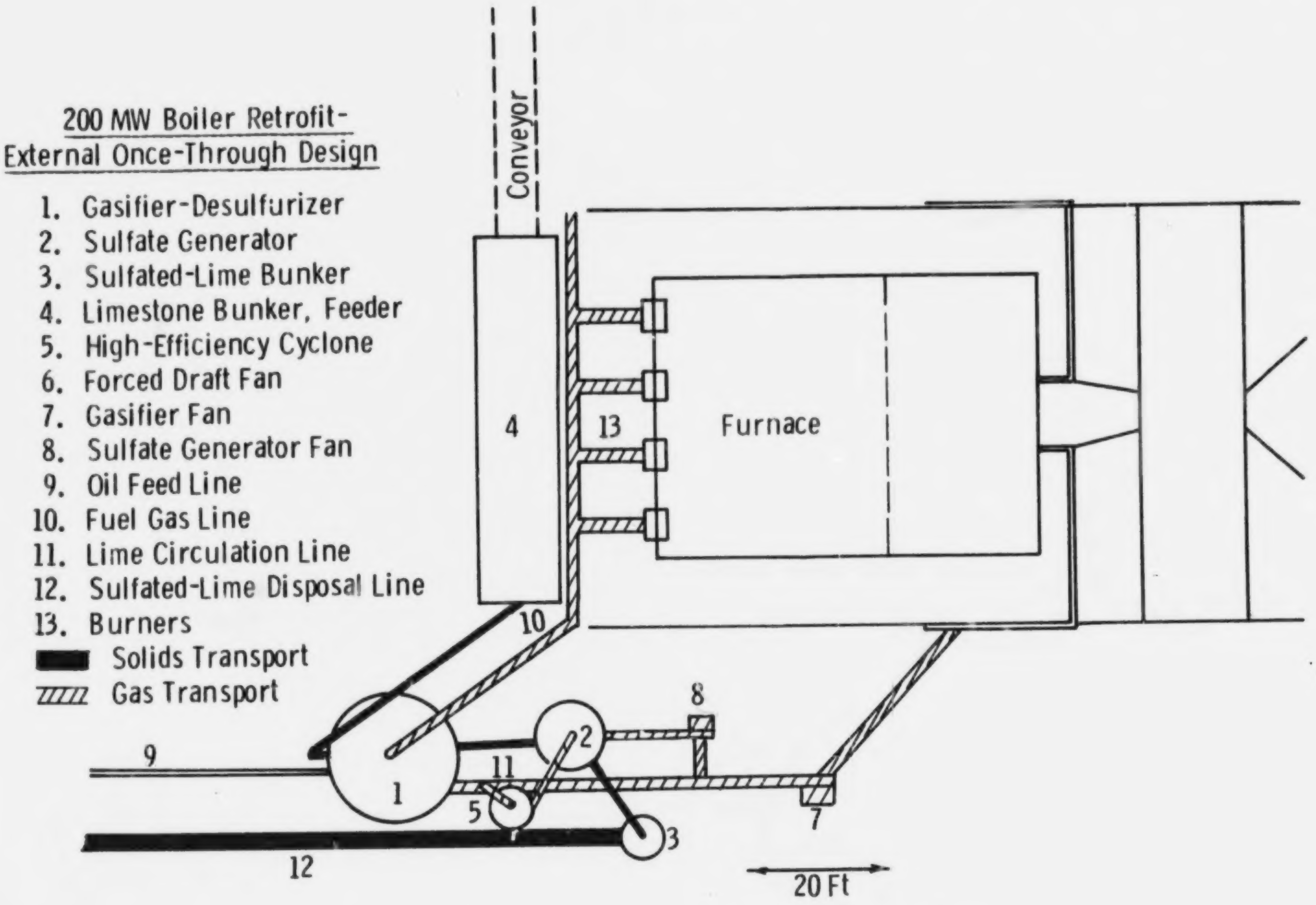


Fig. 3-6—Once-through design-200 MW retrofit

DEVELOPMENT PLANT EQUIPMENT

Gasifier vessel

Regenerator,(1) sulfate generator(2)

Limestone handling

- Limestone bunker(3)
- Conveyors(3)
- Feeders
- Sulfided lime bunker
- Lime transport lines
- Lime disposal line

Storage

- Fuel oil(4)
- Limestone(3)
- Sulfur(1)
- Sulfided lime disposal storage(3)

Particulated clean-up

- Cyclones
- Electrostatic precipitator(3)

Fans, pumps

- Fuel oil(4)
- Gasifier
- Regenerator,(1) sulfate generator(2)
- Pneumatic transport

Burners

Sulfur recovery plant(1)

Piping and ducts

Foundations, structures, electrical, instrumentation, control, experimental equipment

Boiler modifications (removal or alteration of)

- Ash pit(5)
- Pulverizers, pumps, external equipment to be relocated(1, 6)
- Boiler water walls, base of boiler(5)
- Boiler burner ports, structures

---

(1) Regenerative operation only

(2) Once-through operation only

(3) Equipment already present in coal-fired boiler

(4) Equipment already present in oil-fired boiler

(5) Modification with internal design only

(6) Modification with external design only

Boiler Capacity

A development plant with a capacity of 50 to 200 MW may be the best size in that it is representative of the boiler retrofit market and minimizes the scale-up factor. For boiler sizes less than about 50 MW the overall boiler design differs greatly from that of boilers in the 50 to 800 MW range. The 50 to 200 MW capacity boiler may also be more available than a smaller capacity unit. Although the scale-up from the ∿ 1 MW pilot plant unit of Esso (England) to a 50 to 200 MW system is considerable, this represents the minimum scale-up factor which will give meaningful demonstration results within the next five years.

Experimental Program

The demonstration plant should be equipped to monitor the conditions of major streams and vessels to demonstrate the overall performance of the atmospheric pressure fluidized bed gasification-combustion system. The following areas should be emphasized in investigations with regenerative or once-through operation:

- Boiler Performance:

  Steam conditions
  Combustion Characteristics of flame
  Long term erosion/corrosion of boiler and burners

- Gasification/Desulfurization System Performance:

  Performance of temperature control schemes
  Gasifier thermal efficiency
  Fuel and solids distribution in gasifier
  Sulfur recovery-regeneration performance (regenerative mode)
  Sulfate generator performance (once-through mode)
  Solids circulation and control
  Long term erosion/corrosion, material build-up in system

- System Performance:

  Overall system efficiency
  Optimum operating conditions for a given sulfur removal (in terms of limestone make-up rate, bed depth, particle size)
  Pollution control - NO_x, SO_2, particulates, metal retention
  Effects of turn-down on system performance
  Overall system operation.

The experimental program will be determined on the basis of the concept and the availability of the plant for experimental work.

Cost Estimate

Major equipment for a 100 MW demonstration plant (regenerative operation) are estimated in Table 3-4: the cost of instrumentation, control, and piping for the demonstration plant has been taken as a factor twice that expected with a commercial plant, or 100% of the equipment cost rather than 50%, because of the need for flexibility in the demonstration plant functions and the additional costs associated with the experimental nature of the system. Engineering costs are also assumed to be a larger factor with the demonstration plant than with a commercial unit. The regenerative demonstration plant is estimated to cost about $80/kw, while a once-through development plant of 100 MW capacity would cost about $60/kw, or $6,000,000. The major difference in capital cost for the regenerative and once-through operations is associated with the cost of the sulfur recovery plant.

Demonstration Plant Schedule

The schedule for the atmospheric pressure oil gasification/desulfurization development plant program is shown in Figure 3-7.

TABLE 3-4

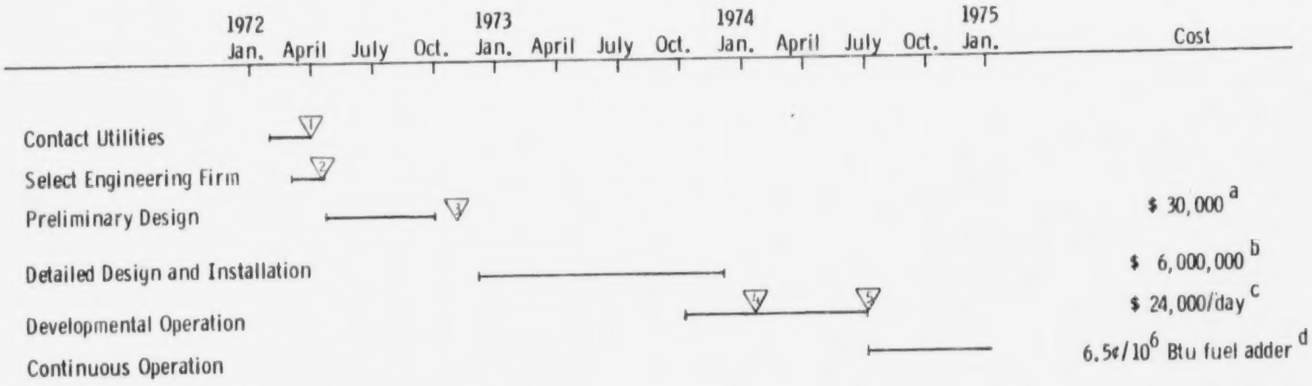
DEVELOPMENT PLANT COST ESTIMATE

ITEM	ESTIMATED COST, $
Gasifier	$ 250,000
Regenerator	35,000
Limestone handling equipment	800,000
Burners	250,000
Particulate control	300,000
Fans	100,000
Boiler modification	300,000
Equipment cost	$2,035,000
Structures, piping, electrical, instrumentation, Control ≈ 100% of equipment	2,035,000
Direct cost	$4,070,000
Engineering, construction ≈ 50% of direct cost	2,035,000
Sulfur recovery plant	2,000,000
Total cost of demonstration plant	$8,105,000

Dwg. 723B675

FIG. 3-7—DEMONSTRATION PLANT SCHEDULE

Atmospheric Oil Gasification/Desulfurization Process for Existing Steam Power Plants



Milestones
- ▽1 Cooperating utility identified
- ▽2 Engineering firm contract signed for preliminary design
- ▽3 Decision on construction; agreement between OAP - utility - W - AE
- ▽4 Shake down complete
- ▽5 Engineering and economic data for future operations complete

Footnotes
- a Cost estimate for preliminary design
- b Assumes complete 100 MW gasifier module, once-through limestone system
- c Maximum cost per day assumes continuous oil feed, O & M for gasifier development, and no power generation
- d Based on conceptual design

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ACKNOWLEDGMENTS

The results, conclusions, and recommendations presented in this report represent the combined work and thought of many persons at Westinghouse, at the Office of Air Programs (OAP), and elsewhere. Westinghouse personnel in many divisions throughout the Corporation have contributed. Our subcontractors -- Erie City, Foster Wheeler, and United Engineers and Constructors -- have added their expertise in boiler and power plant designs. Other OAP contractors have freely shared with us their ideas and the results of their research and development effort. Suppliers of boiler and power plant auxiliary equipment have discussed with us their products and provided cost estimates. Other commercial firms with background in fluidized bed technology have received us, shown us their work, and commented on ours. We have attempted to recollect all these contributions and what is undoubtedly (and regrettably) a partial list of organizations and persons is presented on the following pages.

In particular, however, we want here to express our high regard for and acknowledge the contribution of the personnel at OAP who conceived the overall fluidized bed combustion boiler effort and who have defined, monitored, and supported the efforts of Westinghouse and others on the program. Mr. P. P. Turner, Chief of the Advanced Process Section, has served as project officer on our work. Numerous enlightening and helpful discussions have been held with Mr. Turner; with section members D. Bruce Henschel and Sam Rakes; and with R. P. Hangebrauck, Chief of the Demonstration Projects Branch.

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The following organizations and groups acted as subcontractors:

Foster Wheeler Corporation	● Assisted in evaluating utility fluidized bed boiler design concepts ● Prepared preliminary designs of preferred concepts ● Prepared cost estimates	R. W. Bryers J. D. Shenker R. Zoschak
United Engineers and Constructors Inc.	● Prepared plant layout, operating procedure, and plant cost for a pressurized fluidized bed combined cycle plant	M. Casapis E. Berman W. Craig J. Crowley
Erie City Energy, Division of Zurn Industries	● Compiled industrial boiler market survey ● Assisted in evaluating industrial fluidized bed boiler design concepts ● Prepared preliminary design of preferred concept ● Prepared cost estimate	R. V. Seibel W. Schwinden
Westinghouse Divisions		
Heat Transfer	● Prepared cost estimates for regeneration system vessels and stack gas coolers	R. Giardina S. Jamison
Small Steam and Gas Turbine	● Prepared design and cost estimate for external combustor gas turbine	R. E. Strong B. Hugoson
Power Systems Generation Planning	● Prepared utility market survey ● Provided power system cycle analyses	S. J. Jack R. R. Boyle R. J. Budenholzer H. L. Smith N. Weeks
Computer and Instrumentation	● Prepared control system design and cost for industrial boiler	R. Gessford
Research Laboratories	● Gas Turbine Erosion and Corrosion	W. E. Young

Sturtevant	• Supplied design and cost data for forced and induced draft fans	R. Mercer W. F. Courtney
Large Turbine	• Supplied cost information on steam turbines	

The following vendors supplied information and cost estimates:

Aerodyne Development Corporation	• Supplied design and cost information for high efficiency particulate removal equipment
Petrocarb, Inc.	• Supplied design and cost information for pressurized solids feeding systems
McNally-Pittsburg Manufacturing Corporation	• Supplied design and cost information for solids handling systems
Struthers Nuclear and Process Company	• Submitted proposal for stack gas coolers
Ducon Company, Inc.	• Supplied cost information for particulate removal
Ford, Bacon, and Davis, Inc.	• Supplied cost estimates for sulfur recovery processes
Ralph M. Parsons Company	• Supplied cost estimates for sulfur recovery processes
Wheelabrator Corporation	• Supplied electrostatic precipitator cost
Buell Engineering Company, Inc.	• Supplied electrostatic precipitator cost
Koppers Company, Inc.	• Supplied design information on solids feeding systems
Process Combustion Corp./ Bloom Engineering Co.	• Supplied design and cost information on low Btu gas burner for boiler

The following facilities were visited by Westinghouse personnel:

Argonne National Laboratory
Chicago, Illinois

The Badger Company, Inc.
Boston, Massachusetts

Battelle Memorial Institute
Columbus, Ohio

Bituminous Coal Research
Monroeville, Pennsylvania

Chicago Bridge & Iron Company
Research Center
Plainfield, Illinois

Consolidation Coal Company
Pittsburgh, Pennsylvania

Dorr-Oliver Fluidized Bed
Sludge Incinerator
Liberty, New York

Esso Research and Engineering
Linden, New Jersey

Esso Petroleum Company
United Kingdom

Fluidized Bed Course
University of Birmingham, England

Fuller Company
Catasauqua, Pennsylvania

Kansas Power & Light
Topeka, Kansas

NAPCA/NCB Fluidized Bed Combustion Program
Second Information Exchange Meeting
Hobart House, London
BCURA, Leatherhead
CRE, Cheltenham

National Coal Board
United Kingdom

Pope, Evans & Robbins
Alexandria, Virginia

St. Joseph Lead Company
Monaca, Pennsylvania

Tennessee Valley Authority
Chattanooga, Tennessee

UGI Corporation
Kingston, Pennsylvania

Union Electric
St. Louis, Missouri

U.S. Bureau of Mines
Bruceton, Pennsylvania

U.S. Bureau of Mines
Morgantown, West Virginia

West Virginia University